PB-246 051

ADVANCED WASTE TREATMENT SEMINAR, SESSION IV, COMBINED TREATMENT AND APPLICATIONS, HELD AT SAN FRANCISCO, ON OCTOBER 28-29, 1970

Federal Water Quality Administration

Prepared for:

California State Water Resources Control Board

October 1970

DISTRIBUTED BY:

NTIS

National Technical Information Service
U. S. DEPARTMENT OF COMMERCE

282040

PB 246 051

CLEAN WATER

SESSION IV

COMBINED TREATMENT & APPLICATIONS

ADVANCED WASTE TREATMENT SEMINAR

San Francisco
October 28 & 29
1970

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S Department of Commerce
Springfield, VA. 22151

BIBLIOGRAPHIC DATA SHEET	1. Report No.	2.	3. Recipient's Accession No. PB 246 051
4. Title and Subtitle Advanced Waste Treatment Seminar - Session IV - Combined Treatment & Applications			5. Report Date Oct. 1970
			6.
7. Author(s)			8. Performing Organization Rept. No.
9. Performing Organization Name and Address Advanced ~~Water~~ Waste Treatment Seminar San Francisco, Calif.			10. Project/Task/Work Unit No.
			11. Contract/Grant No.
12. Sponsoring Organization Name and Address California State Water Resources Control Board & Federal Water Quality Administration			13. Type of Report & Period Covered
			14.

15. Supplementary Notes

16. Abstracts This session of the seminar dealt with the treatment of wastes - the physical & chemical treatment specifically nitrogen removal & demineralization of water. Examples of advanced waste treatment ~~at Lake Tahoe~~ at Lake Tahoe are discussed

17. Key Words and Document Analysis. 17a. Descriptors

* Waste treatment
Demineralizing

17b. Identifiers/Open-Ended Terms

*, nitrogen removal
Lake Tahoe

17c. COSATI Field Group

PRICES SUBJECT TO CHANGE

18. Availability Statement	19. Security Class (This Report) UNCLASSIFIED	21. No. of Pages 98
	20. Security Class (This Page UNCLASSIFIED	22. Price 4.75-2.25

ADVANCED WASTE TREATMENT SEMINAR

KABUKI THEATRE
1881 POST STREET
SAN FRANCISCO

OCTOBER 28-29, 1970

JOINTLY SPONSORED BY
CALIFORNIA STATE WATER RESOURCES CONTROL BOARD
AND
PACIFIC SOUTHWEST REGIONAL OFFICE
FEDERAL WATER QUALITY ADMINISTRATION

PROGRAM

OCTOBER 29 - AFTERNOON

SESSION IV COMBINED TREATMENT AND APPLICATION

Moderator: Mr. Frank Middleton

1:15 PM Physical - Chemical Treatment - Jesse M. Cohen

1:45 PM Integration and Management of Advanced Waste Treatment Processes at Lake Tahoe - Russell L. Culp

2:45 PM Nitrogen Removal from Agricultural Waste Water Louis A. Beck

3:15 PM Demineralization by Reverse Osmosis, Electrodialysis, and Ion Exchange - John N. English

3:45 PM Summary: Applications of Advanced Waste Treatment Processes - Mr. Middleton

4:15 PM Closing Remarks - Paul De Falco, Jr.

CONTENT

Physical - Chemical Treatment.....Jesse M. Cohen

Integration of Advanced Wastewater Treatment Processes..... Russell L. Culp

Nitrogen Removal from Agricultural Waste Water.....Louis A. Beck

Mineral Removal by Ion Exchange, Reverse Osmosis and Electrodialysis.....John N. English

The Seminar also included the below sessions:

Session I	Nitrogen Removal
Session II	Phosphorus Removal
Session III	Removal of Solids and Organics

Jesse M. Cohen

I. INTRODUCTION

In recent years it has become apparent that conventional biological treatment may not be the optimum solution to all waste treatment problems. There have been numerous instances wherein the wastewater under consideration contained non-degradable substances or materials which were deleterious to the performance of biological systems. Operating difficulties, sludge handling problems and large land area requirements are intrinsic to biological processes and have led to consideration of alternatives to the activated sludge and trickling filter processes for so-called secondary treatment.

II. DESIRED EFFLUENT QUALITY

First of all, let us define what effluent quality we consider equivalent to secondary effluent. It is fair to say that an effluent BOD of less than 10 mg/l, COD of less than 60 mg/l, and suspended solids of less than 10 mg/l, could be considered good quality secondary effluent. It is clear then that efficient removal of organic material and suspended solids are the major functions of alternative processes for secondary treatment. With the current interest in phosphorus removal for the control of eutrophication we should also consider that function in the development of alternative processes.

III. THE CLARIFICATION-ADSORPTION PROCESS

The most promising combination of unit processes which will produce the desired effluent quality at a reasonable cost appears to be chemical clarification followed by adsorption on activated carbon. A simplified schematic flow diagram of a typical coagulation-adsorption process is shown in Fig. 1.

A. Clarification

Here raw waste water, after screening and grit removal, is treated with a coagulating chemical - lime, iron or aluminum compounds, polyelectrolytes, or a combination thereof. Iron, alum and lime are all excellent phosphorus precipitants. The dosed wastewater is flocculated and clarified and the sludge from the clarification system can be dewatered and disposed of, or in the case of lime, can be recalcined for lime recovery if found to be economical. Experience at Lake Tahoe[1] on tertiary treatment with lime has shown that no significant saving in chemical cost is achieved by recalcining the lime sludge. However, sludge disposal costs are reduced because only a fraction of the lime sludge must be bled off to disposal to limit the buildup of inerts. Recalcined and

---

Presented at the Advanced Waste Treatment Seminar, Pacific Southwest Region, FWQA, San Francisco, California, October 28-29, 1970.

reused lime has proven to be as effective as commercial lime in clarification and phosphorus removal. Recovery schemes for iron and aluminum sludges have not yet been developed and so the use of these coagulants will probably be limited to smaller installations where coagulant recovery would not be economical. Polyelectrolytes have been used on plant scale to enhance primary clarification but they have no phosphorus removal capability and thus their use as primary coagulants in new physical-chemical processes will probably be limited. They do have the advantage of minimizing increased sludge production.

Clarified waste can then be filtered - preferably by dual-media filtration - or applied directly to carbon adsorption systems.

B. Carbon Adsorption

There are two general types of carbon adsorption systems, granular carbon systems and powdered carbon systems. Of these granular carbon holds the most promise but powdered carbon systems are also being investigated and may find some application. In the common granular carbon systems being considered, the clarified wastewater is passed through a bed of granular activated carbon particles, usually 8x30 or 12x40 mesh, where organic molecules are adsorbed on the carbon surfaces. At such time as the capacity of the carbon to adsorb additional organic materials is decreased to such an extent that the effluent COD increases to a predetermined limit the carbon bed is removed from service. The spent carbon is transferred to a thermal regeneration system where the adsorbed organics are volatilized and driven off the carbon surface, and the adsorption capacity of the carbon is thus restored.

In powdered carbon systems clarified wastewater is contacted with a slurry of powdered activated carbon. The carbon is separated by flocculation with polymers, followed by clarification and filtration. Powdered carbon regeneration systems are in the development stage but several appear promising. Until the regeneration of powdered carbon has been successfully demonstrated granular carbon will be the adsorbant of choice.

C. Filtration

A dual-media filtration system could be utilized either before or after the adsorption system for nearly complete removal of suspended solids. The filtration system must follow powdered carbon adsorption systems for complete removal of carbon particles.

IV. PERFORMANCE OF PHYSICAL-CHEMICAL SYSTEMS

A. Pilot Plant Study[2]

Only limited data have been generated in studying the combination of processes discussed here. Under an FWPCA contract the FMC Corporation performed a pilot-scale investigation of the coagulation-adsorption process for treatment of primary effluent. One objective of this study was a parallel comparison of adsorption in downflow packed beds and upflow fluidized beds. Figure 2 shows a schematic of the clarification system used. Primary effluent was fed to the system at a rate of 5.5 gpm.

An average dose of 170 mg/l $FeCl_3$ was added, followed by two minutes of mixing at 1500 rpm and 15 minutes of flocculation at 18 rpm. Clarification was accomplished in a one-hour detention time upflow clarifier. The clarifier effluent was filtered by an anthracite-sand dual-media filter.

Clarified waste was applied to parallel sets of carbon columns, one upflow fluidized bed and the other downflow packed bed as shown in Figure 3. The rate was 5 gpm/sf in each column and the empty bed contact time was 64 minutes. Figures 4, 5, 6 and 7 show the results of the study. Considerable organic removal took place in the clarification system. Final effluent TOC was around 5 mg/l, BOD around 5 mg/l, turbidity around 1 JTU and phosphate 2-3 mg/l as PO_4 (around 1 mg/l as P).

B. Comparison of Carbon Adsorption and Activated Sludge[3]

A comparison of carbon adsorption and activated sludge for the removal of organic material from primary effluent was made over a six-week period at Lebanon, Ohio. Settled and filtered primary effluent was applied to granular carbon columns with a contact time of 37 minutes. The carbon column effluents were compared with sand-filtered secondary effluent from the 1 MGD activated sludge plant at Lebanon. Table 2 shows that carbon adsorption was more efficient in the removal of organics than activated sludge.

C. Development of a Full-Scale Installation

Several studies have been made utilizing polymer addition to existing primary plants followed by small-scale pilot carbon adsorption. One such study was done at the 10 MGD primary facility at Rocky River, Ohio[4]. An anionic polymer was added to the existing primary clarifier at a dosage of 0.3 mg/l, and a side stream of

clarified effluent was applied to small carbon columns for a period of about one month. The summary of the data is shown in Table 3. It can be seen that, even with less than optimum clarification, effluent comparable to good secondary effluent was produced with 33 minutes carbon contact time. On the basis of this preliminary work the City of Rocky River applied for and was awarded a Research and Development Grant from FWPCA to help support a full-scale investigation of the clarification-adsorption process for secondary treatment.

One of the principal motivations for the city to use physical-chemical treatment is shown in Figure 8. The installation of conventional activated sludge facilities would necessitate the condemnation of a considerable area of very expensive property, whereas a carbon adsorption system could easily fit into the existing site. Testing has been undertaken to determine what coagulant or combination of chemicals will be used in the clarification system. If phosphorus removal is required an inorganic coagulant will necessarily be the choice. Figure 9 shows the schematic flow diagram of the carbon adsorption system to be constructed at Rocky River. In Table 4 the last column shows the estimated treatment costs at Rocky River. The total cost including amortization is estimated at 10¢/1000 gal. The makeup carbon cost is based on a carbon exhaustion rate of 500 lb/MG, and 5% carbon loss per regeneration cycle. The chemical cost item is for 0.3 mg/l of anionic polymer which may or may not be used. The capital cost of the proposed installation is estimated at $1.6 million.

There is other work going on in the investigation of such systems and before too many years we will have the answers to some of the as yet unresolved questions. The major question is, of course, how much does it cost? When some full-scale systems have been in operation for a period of several years this big question can be resolved.

References

1. Smith, C. E., "Recovery of Coagulant, Nitrogen Removal, and Carbon Regeneration in Waste Water Reclamation," Final Report, FWPCA Grant WPD-85(June 1967).

2. Weber, W. J. Jr., Hopkins, C. B., & Bloom, R. Jr., "Physiochemical Treatment of Wastewater," Presented at the 42nd Ann. Conf. of the Water Pollution Control Federation, Dallas, Texas, Oct. 1969.

3. Schwartz, W. A., Internal Report, AWTRL, FWPCA (1969).

4. Rizzo, J. L., and Schade, R. E., "Secondary Treatment with Granular Activated Carbon," Water & Sew. Wks., 116, No. 8, pp.307 (1967).

TABLE 1

DESIRED EFFLUENT QUALITY

BOD	10 MG/L
COD	60 MG/L
SS	10 MG/L
P	1 MG/L

TABLE 2

COMPARISON OF PHYSICAL-CHEMICAL TREATMENT
TO BIOLOGICAL TREATMENT
AT LEBANON (8/21-10/3/69)

(Effluent quality, mg/L)

	TOC				
	Total	Soluble	COD	Color	Turbidity
PRIMARY EFFLUENT	87.0	32.5	309	--	--
Settled	68.5	30.2	251	53	57
Dual-Media Filtered	49.0	26.7	186	45	23
CARBON EFFLUENT-S	11.7	7.3	57	28	13
CARBON EFFLUENT-F	8.3	5.5	48	26	6.6
SAND-FILTERED SECONDARY EFFLUENT	25.6	16.1	64	14	8.4

S - feed of settled primary effluent

F - feed of dual-media filtered primary effluent

TABLE 3

Rocky River Waste Treatment Plant
Clarification/Carbon Process

	Raw Water	Clarified Water	Carbon Contact Time, Minutes				Percent Removed
			4.7	14	23.4	32.6	
Suspended Solids, mg/l	107	65	31	13	15	7	93.3
BOD, mg/l	118	57	27	21	11	8	93.3
COD, mg/l	235	177	117	67	50	44	81.3
TOC, mg/l	52	53	33	18	15	13	75

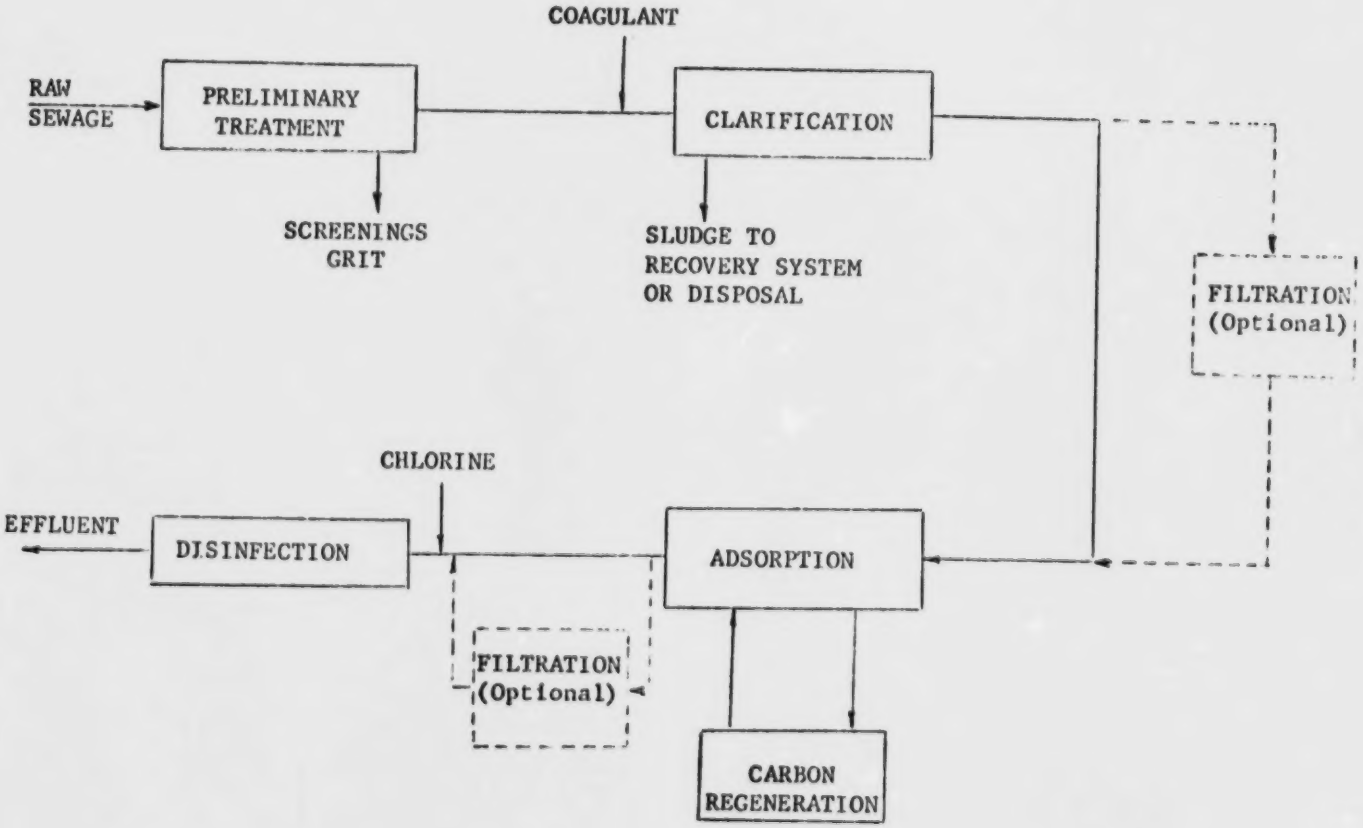
TABLE 4

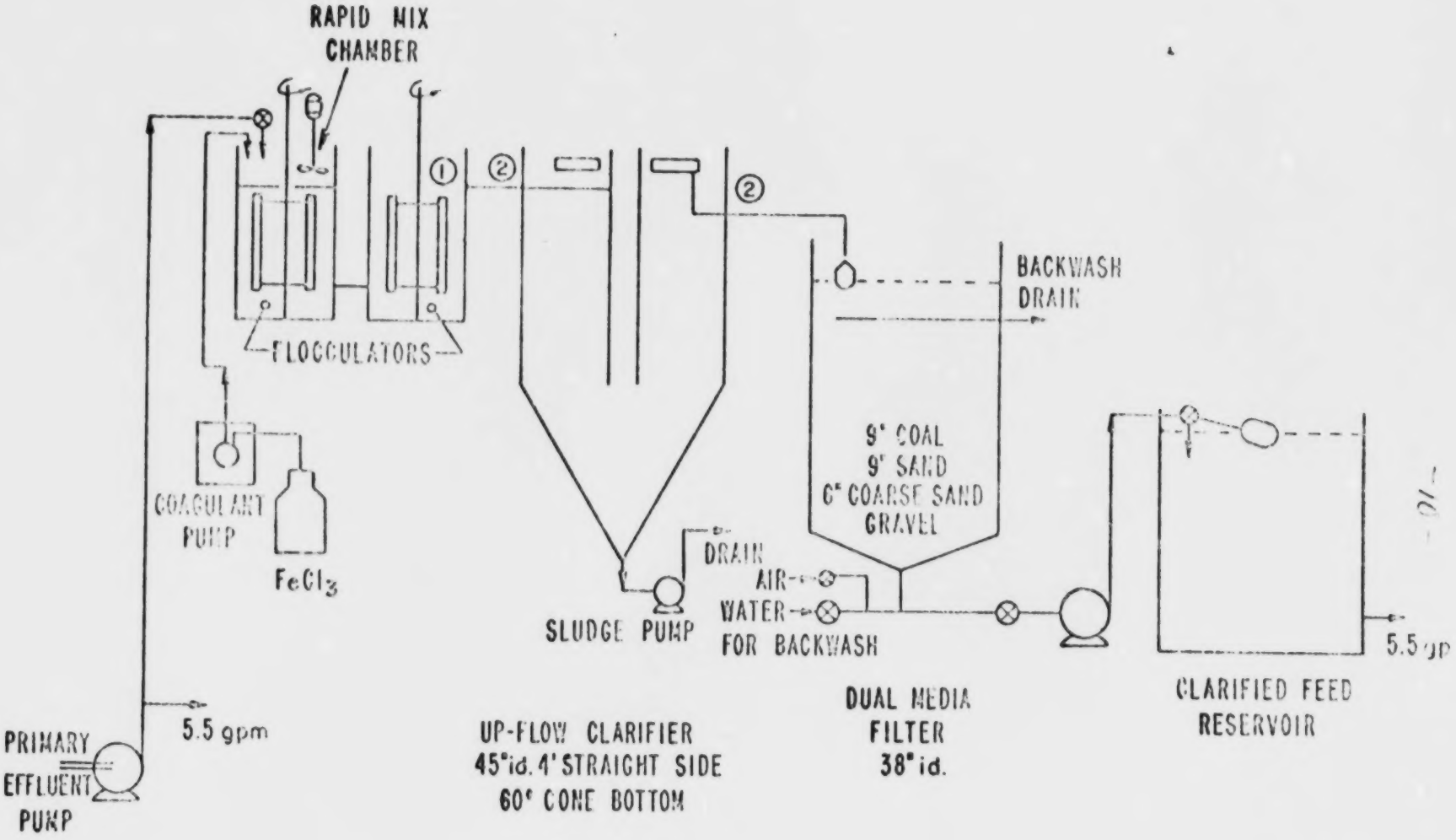
Capital And Operating Costs
Granular Carbon Adsorption

	Pittsburgh Activated Carbon Co.	Lake Tahoe	Pomona	Rocky River
Capacity, mgd	10	7.5	10	10
Investment ($1,000)	1,489	1,306	1,670	1,600
Operating Cost, (¢/1000 gal.)				
Carbon	1.20	1.18	1.10	0.69
Fuel	0.11	----	0.25	0.12
Chemicals	----	0.99	----	3.80
Power	0.85	0.75	0.85	0.55
Labor	0.74	0.40	1.50	1.10
Overhead	0.27	----	----	----
Amortization	3.07 (20 yrs)	3.53 (20 yrs)	4.10 (15 yrs)	3.23 (20 yrs)
Maintenance	0.63	0.33	0.50	0.55
Total Operating Cost	6.87	7.18	8.30	10.04

FIGURE 1

FLOW DIAGRAM OF A PHYSICAL-CHEMICAL TREATMENT SYSTEM





FLOW DIAGRAM OF CLARIFICATION SYSTEM

FIGURE 2

EXPERIMENTAL SET-UP FOR 24 FT. CARBON BEDS

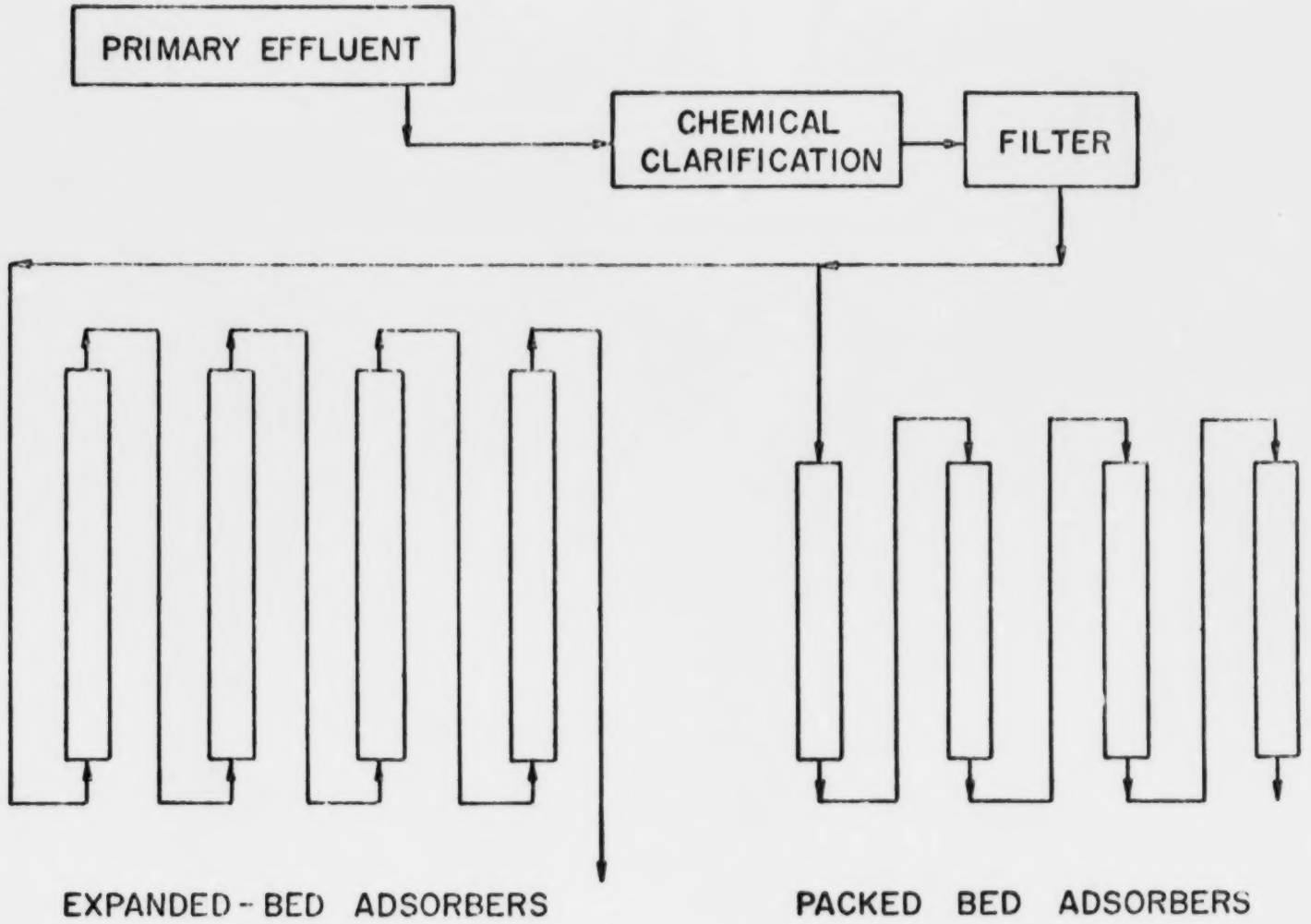
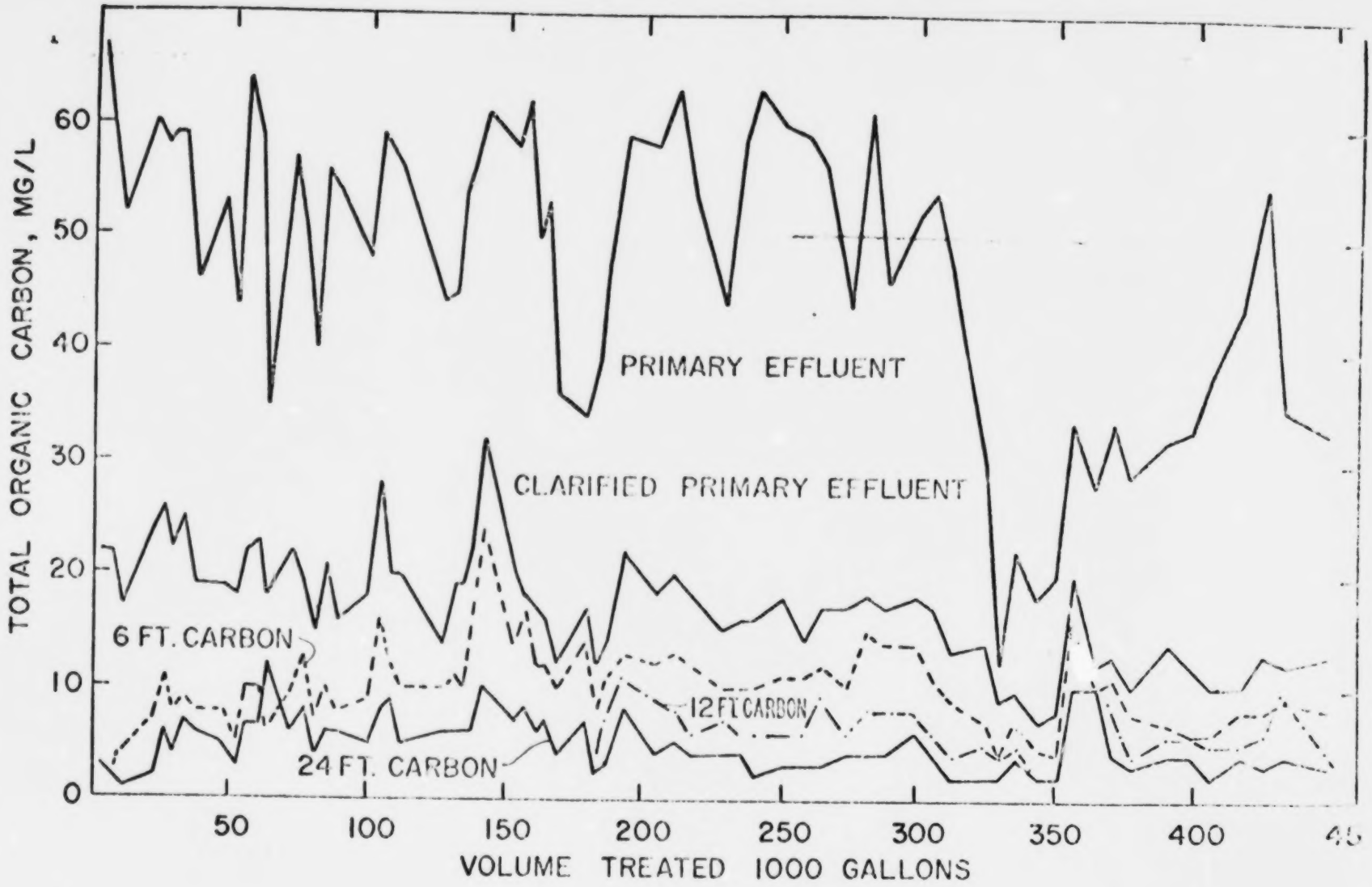


FIGURE 3



TREATMENT OF PRIMARY EFFLUENT BY CLARIFICATION AND ACTIVATED CARBON IN EXPANDED BEDS FIGURE 4

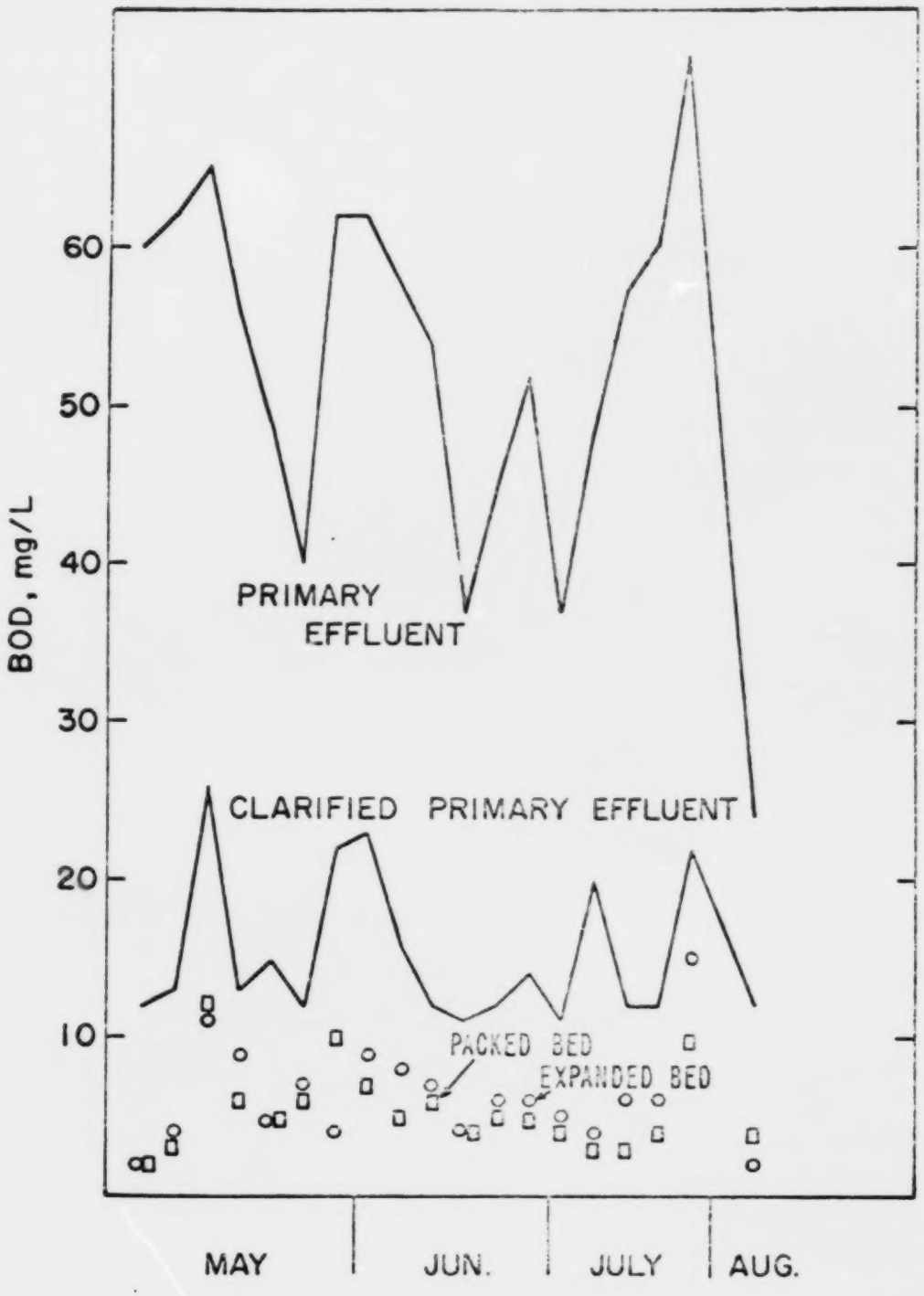


FIGURE 5. REMOVAL OF BOD FROM PRIMARY EFFLUENT BY CHEMICAL CLARIFICATION AND 24 FT. ACTIVATED CARBON

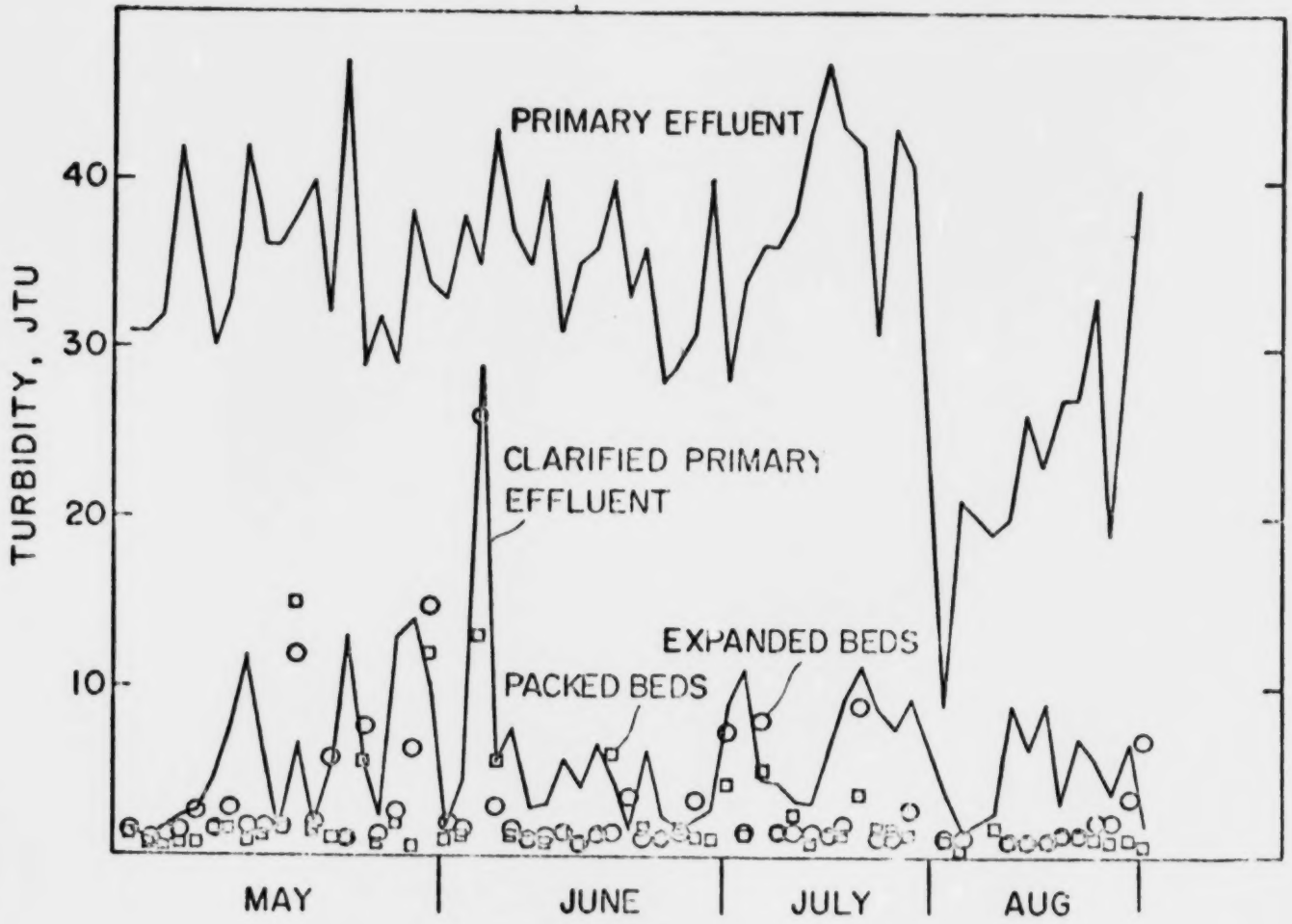
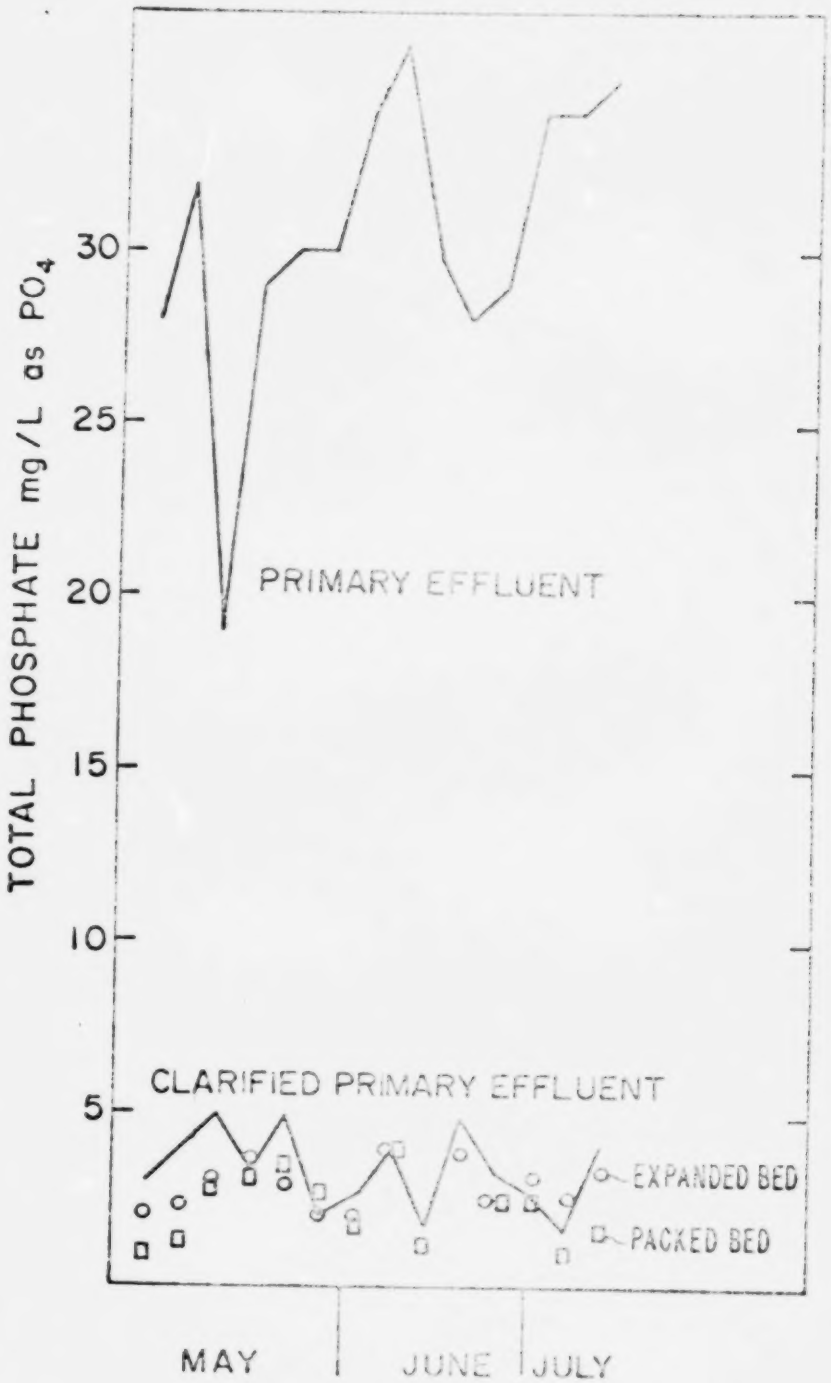
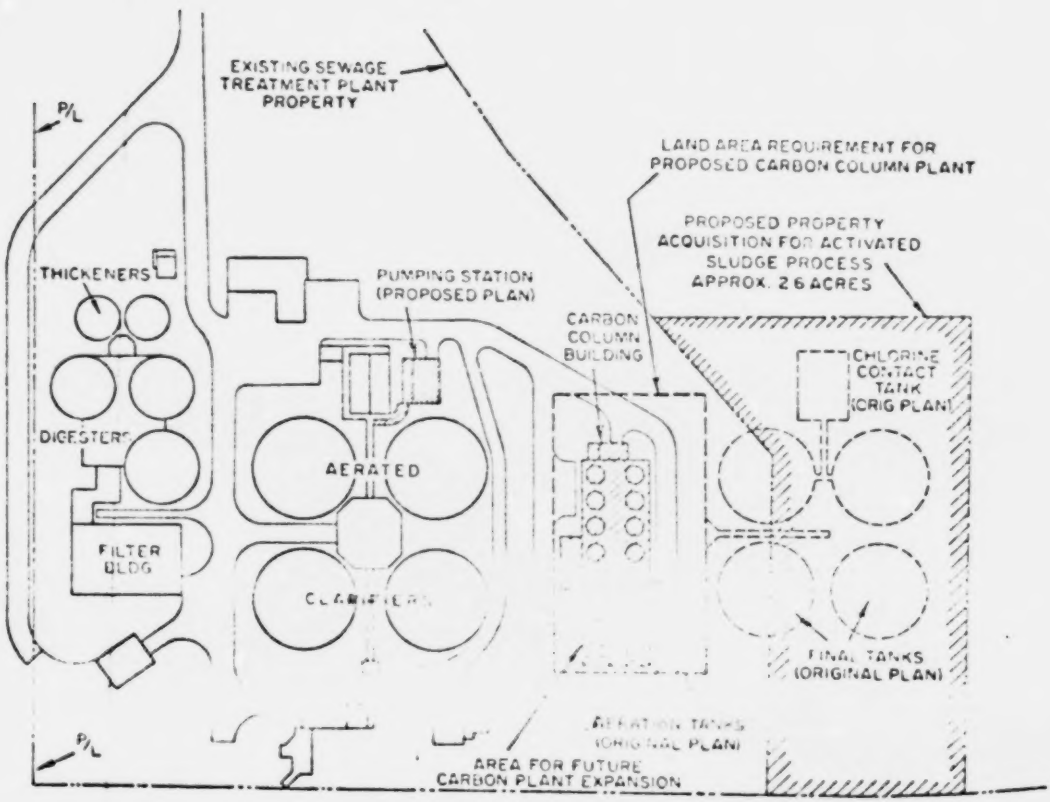


FIGURE 6. REMOVAL OF TURBIDITY FROM PRIMARY EFFLUENT BY CHEMICAL CLARIFICATION AND ACTIVATED CARBON BEDS



REMOVAL OF PHOSPHATE FROM PRIMARY EFFLUENT BY CHEMICAL CLARIFICATION AND 24 FT. ACTIVATED CARBON FIGURE 7.



Area comparison for clarification/carbon process vs activated sludge plant at Rocky River.

FIGURE 8

Reproduced from best available copy.

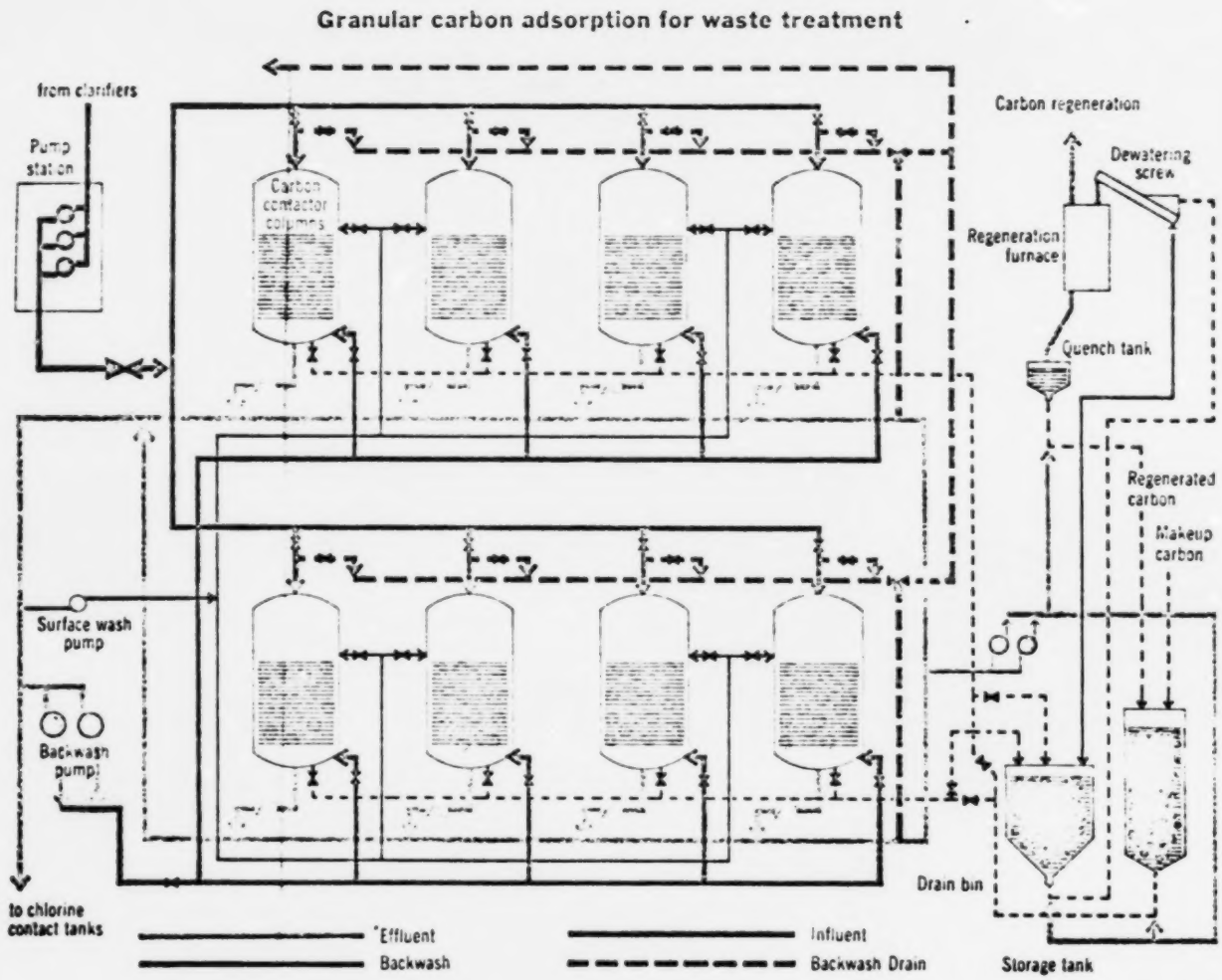

Granular carbon adsorption for waste treatment
from clarifiers
Pump station
Carbon contactor columns
Carbon regeneration
Dewatering screw
Regeneration furnace
Quench tank
Regenerated carbon
Makeup carbon
Surface wash pump
Backwash pump
Drain bin
to chlorine contact tanks
Effluent
Backwash
Influent
Backwash Drain
Storage tank

FIGURE 9

INTEGRATION OF ADVANCED WASTEWATER TREATMENT PROCESSES*

By Russell L. Culp**

Introduction. Previous papers presented at this seminar have described various unit processes which are available for advanced wastewater treatment. This discussion is concerned principally with considerations involved in selecting and sequencing unit processes to produce water which will meet various quality criteria for wastewater disposal or for various municipal, industrial, agricultural, or recreational uses of reclaimed water. In addition, brief presentations will be made regarding operation and cost experiences in 2-1/2 years of full-scale advanced wastewater treatment at South Tahoe.

Factors Affecting Selection of Processes. Some of the factors affecting selection and sequencing of processes are shown in Table 1.

Table 1

Partial Check List of Factors Affecting Selection and Sequencing of Unit Processes.

1. Effluent quality required
   a. present
   b. possible future
2. Reliability and simplicity
   a. unit processes
      (1) effects of changes in flow, temperature, and applied water quality
      (2) required operator skill and attention
      (3) process sequence interrelationships
         (a) effects of secondary treatment upsets
3. Status of process development
   a. laboratory experiments
   b. pilot plant
   c. operating full-scale plants
      (1) design data and equipment specifications
      (2) actual operating experience
      (3) firm cost data
4. Requirements for ultimate disposal of sludge.
   a. land disposal
      (1) raw sludge
      (2) digested sludge
      (3) dewatered sludge
   b. incineration

* Presented to the FWQA Advanced Waste Treatment Seminar

**General Manager, South Tahoe Public Utility District, South Lake Tahoe, California

(1) sludge dewatering characteristics
  (a) primary sludge
  (b) waste activated sludge
  (c) chemical sludge
  (d) organic-chemical sludge mixtures
(2) sludge conditioning
  (a) digestion
  (b) heat treatment

5. Process compatibility.
  a. optimum points for chemical application
  b. optimum pH for each unit process
  c. recycling of process water and waste streams
    (1) centrate
    (2) filtrate
    (3) scrubber water
    (4) filter backwash water
    (5) carbon rinse water
    (6) miscellaneous
6. Process flexibility.
  a. alternate modes of operation
  b. standby units
  c. provisions for servicing and repair
  d. fail safe features
    (1) process monitoring
    (2) restricted bypass arrangements
    (3) alarms
7. Costs.
  a. capital
  b. operating
  c. plant capacity or size of units
    (1) powdered vs granular carbon
    (2) gravity vs pressure units
    (3) materials of construction
    (4) throwaway vs reuse of chemicals
8. Awkward process combinations. (Carrying out two or more processes simultaneously in a single treatment unit.)
  a. general problems
    (1) loss of individual process control and optimization
    (2) loss of flexibility
    (3) loss of reliability
  b. mixing, flocculating, settling, and internal solids recycling in a single basin (solids contact basin)
  c. chemical-primary treatment
  d. chemical-activated sludge treatment
  e. carbon contactor-filter
  f. sludge incineration-lime recalcining in a single furnace
9. Future reuse for potable supply, special considerations.
  a. holding time
  b. double chlorination with intermediate contact time
  c. exclusion of certain industrial wastes
  d. makeup water low in total solids, or demineralization of portion of wastewater

1. Effluent Quality Required.

The first consideration in the selection of unit processes is the degree of treatment required at present and that which may be required in the future. Once the required water quality is established, then unit processes must be selected to accomplish the desired result. Obviously, for almost every situation, there are a number of combinations of unit processes which may satisfy the needs.

2. Reliability.

The simplicity and reliability of unit processes and sequences thereof are top priority factors. While simplicity and reliability are not synonymous, they are very closely related. The simpler the process is to operate and control, the more likely it is to operate successfully and continuously. Regardless of how highly efficient a process may be when operating at its very best under conditions rigidly controlled by experts, if it is sensitive to minor changes in flow, temperature, or applied water quality, or if it requires constant attention and adjustment, then it is not likely to operate as intended all of the time in practical plant application.

The operation of conventional secondary treatment plants is often very erratic. The design of advanced wastewater treatment processes which follow secondary treatment processes must take this fact into account. They must be able to accept these variations in secondary plant performance and still produce an effluent of uniform quality.

For example, it is good to follow activated sludge with effective chemical coagulation and settling or mixed media filtration in order to capture biological solids which often escape plain sedimentation, or with granular carbon to equalize variations in BOD or COD content.

3. Status of Process Development.

The state of development of a process, particularly the number of full-scale plants from which engineering design and equipment information, actual operating results, and cost data can be obtained, has a great influence on the degree of reliability which may be expected of it. Obviously, if new approaches are not tried on a plant-scale, little progress will be made. However, engineers and clients embarking in new areas should be aware of the degree of uncertainlty involved. The acceptable risk level must be determined individually for each project.

The first tests of any new treatment method are usually made in the laboratory. The next step is pilot plant operation on a continuous flow basis. Pilot plant work gives a good insight into the results to be expected from liquid processing. Unfortunately, it does not always adequately evaluate sludge problems or problems associated with recycled process streams or return solids. Only a few advanced wastewater treatment processes have been tested on a plant scale. It is only at plant scale that many engineering design problems and the practical operation problems are faced for the first time. The transformation of pilot plant data gathered by biologists, engineers, chemists, and

laboratory technicians into workable engineering designs is a difficult and demanding task, and one not to be taken lightly or for granted. It is here, in the hands of the sanitary engineering designer, that the project either fails or succeeds. The success depends not only on how well he selects and combines the unit processes, but more importantly how well he selects and specifies the performance of equipment, and how well the executes the excruciating details of design. Actually, the development of this engineering design know-how is the great gap in the application of new and better methods for wastewater treatment. There is a great backlog and wealth of information and scientific research which cannot be used until it is transformed into actual working plants. This transformation requires the design of new structures and equipment, and the provision of sufficient safety and flexibility in the design to allow for any contingencies which may not be apparent at pilot scale.

Because the problems involved in going from pilot to full scale are difficult, it is sensible to take full advantage of all prior knowledge. There are many unknowns still remaining upon completion of pilot scale experiments which become readily discernable upon completion and operation of the full scale plant. For maximum reliability, the designers of future plants must take advantage of the prior art, modifying proven designs as necessary to meet different conditions as to size and other design variables.

4. Requirements for Ultimate Disposal of Sludge.

Of major importance in process selection are the circumstances related to the handling and ultimate disposal of the sludge produced. In locations where there are available remote, large areas of land, almost any kind of sludge, wet or dry, stable or decomposing, can be, and is, disposed of by hauling or pumping to these land disposal sites. In many places this crude method for sludge disposal probably will not be tolerated indefinitely, but might suffice for the time being. Ocean disposal of sludge has long been an easy way to evade the knotty problems involved in proper sludge disposal. However, the nuisances which have been created and the damages wrought to beaches and coastal waters have aroused the public to the point where this method is now in almost universal public disfavor.

All of the more acceptable methods for sludge disposal involve dewatering of the sludge. As pointed out previously, the ease with which sludge may be dewatered is a prime factor in unit process selection. There are many alternate ways to process the liquid component of wastewater to secure the desired result at about equal costs, but there are very few ways to satisfactorily and economically dewater sludge. In sludge from certain wastewaters, dewatering of mixtures of organic-chemical sludges may be satisfactory, but care must be taken to check this out before designing a full scale plant. Favorable pilot plant test results are an important prerequisite.

Another approach which has been used successfully is to keep entirely separate all organic sludge from all chemical sludges. Then, conventional equipment used for either of these types of sludge can be installed. Pilot tests are still highly desirable, even with this approach.

Heat treatment or anaerobic digestion of organic-chemical sludges may

condition these mixtures to make possible use of a wide range of dewatering equipment. Chemical conditioners may accomplish this same end, but dosages must be determined to ascertain economic feasibility.

5. Process Compatibility.

Another consideration in process selection is the compatibility of the process with other unit processes used in the overall treatment scheme. The possible effects of waste streams or recycled solids is very important.

The optimum pH for various unit processes is a factor which influences their compatibility, and my affect placement of processes in the plant flow sequence. There are many examples of this point. Various chemical coagulants operate best within certain pH ranges. In this case, there may be a choice, either change the pH or change the chemical used. Also it is possible to broaden the effective pH range of chemicals by use of coagulant aids, such as activated silica or polymers. Biological processes may be adversely affected by pH values above 9.3, although activated sludge cultures have great buffer capacity and it may be possible to allow the pH of the influent to activated sludge basins to exceed 9.3 so long as the pH in the aeration tank itself does not exceed this value. Phosphorus removal by lime requires raising the pH to values of 9 to 11. Ammonia stripping also is most efficient at pH values of 10.5 to 11. Calcium reaches minimum solubility at a pH near 9.3 Carbon adsorption of organics is best at low pH values, and is very poor at pH values above 9.0. Chlorination is more effective at low pH values, at low turbidities, and in waters with low chlorine demand. Granular activated carbon adsorption is favored by good clarification of the applied water. Granular activated carbon will adsorb chlorine. With information on process interrelationships and pH effects, then, a logical process sequence for maximum reliability and effluent quality would be biological treatment followed by advanced treatment, or the so-called tertiary sequence. Although the costs may exceed those of approaches combining certain steps, there are at least three advantages to this arrangement which must be given due consideration: the pH is favorable for biological treatment; the organic sludges are kept separate from the chemical sludges which usually minimizes sludge dewatering problems; and the polyphosphates in the raw wastewater are converted to orthophosphate before chemical coagulants are applied, thus avoiding potential interference with chemical coagulation.

In considering compatibility, attention must be given to the point in the overall process at which centrate, filtrate, backwash water, scrubber water and other plant process water and recycle streams are introduced. It is wise to provide more than one point of return for some of these streams, particularly to take care of those times when certain plant units are temporarily out of service.

6. Process Flexibility.

The capability should be provided in advanced wastewater treatment plants for as many alternate modes of operation as possible. Each basin and each piece of equipment must be taken out of service at same time for servicing or repair, and provisions must be made in the design to operate the plant under these conditions. In many cases, duplicate or standby units may be provided. The plant should include as many fail-safe features as possible, including

equipment for process monitoring and alarms to signal plant failures. By-passes around individual units are essential, but no bypass should be provided which will allow water of unacceptable quality to leave the plant.

7. Costs.

Capital and operating costs are important factors in process selection. It is obviously false economy to use a process which will not do the job merely because it is cheap to install and operate. If the designer has a choice between a proven process with a high degree of reliability and a process which may be cheaper, but is of unknown or questionable reliability, then prudence is on the side of the more costly but workable process.

8. Awkward Process Combinations.

On special word of caution is appropriate with regard to the attractive and tempting possibility of combining various processes into a single process or into a single structure. By reducing the total number of structures, there are obvious savings in construction costs. Actually, this initial cost saving is usually the only advantage of combining processes, and there may be many disadvantages. The history of water and wastewater treatment is replete with failures to satisfactorily combine two or more processes. Yet, some of these mistakes are perpetuated in present day designs, and new unworkable combinations may be devised. This is not to say that some of these efforts may not end in success, but only to inject a word of caution and to point out some of the inherent weaknesses in making one process out of two or more. First of all, there is invariably a loss of control or flexibility in operation by combining two processes. A change which benefits one of the processes may interfere with or reduce the efficiency of the other. Also, in plant operation, it is much more difficult to relate cause and effect, which is necessary to make treatment adjustments, because of the greater complexity introduced by the combination. There are many specific examples which can be presented.

Upflow solids contact basins of the sludge blanket type combine the four processes of rapid mixing, flocculation, settling, and solids recycling in a single unit. They have been applied successfully to treatment of a constant flow of waters (principally well waters) which are of constant physical and chemical quality, at a considerable savings over the use of four separate operations. However, these sludge blanket basins are often not stable when operated at varying flow rates, or with waters which have significant variations in chemical composition or physical characteristics, such as temperature or organic content. It may be difficult to change the point of chemical addition in this type of basin if it becomes necessary. Also with changes in flow rate, chemical composition, or physical character of the applied water, or with the accumulation of organics in the sludge blanket, the chemical dosages required to maintain a sludge blanket of the proper density change so frequently and rapidly as to present an impossible operating condition, and the basin fails to function as a settling device. Wastewaters possess properties which make operation of sludge blanket clarifiers difficult. Some plant experiences indicate that this type of equipment can be used satisfactorily for wastewater treatment if the overflow rates are the same as for a conventional horizontal flow clarifier, and if the sludge is removed from the basin as rapidly as it reaches the bottom of the tank. Separate rapid mixing, flocculation, and settling basins with provisions for external sludge recirculation appear to be

more satisfactory than solids contact basins for treatment of wastewater. Mutual interference is avoided, and complete flexibility of operation and control is afforded.

Difficulties and high costs may be involved in dewatering organic-chemical sludge mixtures in some wastewaters. Combining high lime or alum treatment with primary settling with the elimination of a separate chemical clarifier, or even going further and eliminating biological treatment, offers potential savings which are quite substantial. The possibilities certainly warrant investigation in each instance to see if there is a sludge dewatering problem, and, if so, if there is an economical solution to it. This should be done on a pilot plant scale before undertaking the building of a full scale plant.

Combining chemical treatment with activated sludge treatment for phosphorus removal also presents some practical difficulties at full plant scale which should not be overlooked, even though there may be ways to solve the problems. Addition of lime or alum to aeration tanks secures excellent phosphorus removal. The difficulty is that the inert chemical sludge accumulates and selectively displaces the culture of activated sludge organisms over a period of a few weeks, and normal biological treatment ceases. Because of the differences in specific gravity between the chemical and biological sludges it may be possible to separate them and return or waste them as necessary to control the process, but this remains to be demonstrated on a plant scale.

Combination downflow carbon contactor-filters are used successfully in water treatment plant practice where there are low suspended solids in the applied water. However, the carbon contactor-filter is a surface filter, and not an in-depth filter, and as such is subject to the same limitations as any other surface filter in filtering wastewater. The use of granular carbon as a contactor-filter has all of the same disadvantages of other surface filters plus the additional one of being a much deeper bed, which is more difficult to properly clean by backwashing than a shallower bed. Also, the use of carbon as a filter to remove suspended and colloidal matter from the water is likely to blind the carbon pore openings and require removal of the carbon from the bed before its full adsorptive capacity is exhausted.

Flocculation and settling or organics and lime sludge in the same basin poses another problem if the lime is to be recalcined and reused. Calcium carbonate has been recalcined and organics incinerated successfully in a single furnace. However, this means that the recalcined lime contains considerable inert ash. There are two choices, the ash can be recycled with the recalcined lime, or some means can be devised to separate the lime from the ash. If the ash is recycled with the lime it becomes wet again and there are extra costs for fuel for repeated drying of the ash on subsequent cycles. It is necessary to waste about 25 to 35 percent of the recalcined lime on each cycle of use to rid the system of phosphates, and a similar portion of the recycled ash would automatically be wasted in the same operation. This would be an acceptable way to operate, and the decision as to whether to settle and incinerate organics and chemical sludge together or separately can be resolved by a rigid cost analysis including, of course, all capital and operating costs.

Whether to attempt to separate the ash from the reclaimed lime as opposed to recycling the ash with the lime also is strictly a matter of comparative costs, assuming that either a wet separation or a dry separation can be made successfully.

These examples should serve to illustrate some possible problems which may be encountered when attempting to combine two or more unit processess into a single unit. Combining processes usually complicates operation and control and decreases reliability, while the designer should be seeking to simplify operation and control, provide maximum flexibility of operation, and insure greater reliability. There is no point in combining two perfectly good unit processes into one poor one. Because of the potential savings, efforts to determine workable combinations should certainly continue.

One way to reduce capital costs for construction of advanced wastewater treatment plants is to introduce storage of wastewater at some point in the system. This should be done after the water has received sufficient treatment so as not to creat a nuisance on standing. Sufficient storage should be provided to allow the sections of the plant which follow the storage to operate at the average rate for the design maximum day rather than at the peak hourly rate for the maximum day. A rough rule of thumb for separate sanitary sewer systems is to build at least 1.5 to 2.0 million gallons of storage for each 10 mgd of plant capacity. The exact requirements vary for each individual collection system, of course, and the rule of the thumb does not apply to systems receiving storm water as well as sanitary sewage.

9. Special Factors Associated With Future Water Reuse for Potable Supply.

In the future, when situations develop where the only possible solution to expecially severe public water supply shortages is to recycle domestic wastewater, there will be special considerations of reclaimed water quality related to public health which will affect unit process selection, as well as the combination of these processes which make up the total water reclamation plant.

Present concern is not so much <u>can</u> reclaimed wastewater be used as a source of drinking water supply, but rather <u>should</u> it be so used. Indications are that water quality suitable for direct potable use can be produced from wastewater under certain favorable conditions when there is sufficient justification does not now exist in the United States.

At Windhoek, South Africa, which is located in an extremely watershort area, about one-third of the municipal water supply is now obtained from reclaimed wastewater.

If reclaimed water had to be used for potable purposes today, there are several additions and modifications which could be made to increase the safety of the water from a public health standpoint. For example, at Windhoek, maturation, or holding ponds, are a part of the process. Holding water for a period of several days allows the natural die-away of bacteria, virus, and other disease-producing organisms to occur. Perhaps the most effective barrier to possible disease transmission from water reuse would be double chlorination with time for intermediate contact with the reclaimed water. It is only reason-

able to fully utilize this tool in processing wastewater, because, in the final analysis, all public water supplies now obtained from surface sources are dependent upon adequate chlorination as the principal means of protecting the public health.

The above discussion is based on the premise that only about one-third of the total supply would be obtained from this source. With a completely closed loop and with almost all of the supply coming from recycled water, there is a problem with a buildup of dissolved salts which can only be removed by demineralization of a portion of the water.

Processes Selected At South Tahoe.

On the basis of pilot plant tests and some full scale plant experience in 1965 at Tahoe, and in view of the criteria previously presented in this paper, the processes incorporated at the South Tahoe plant are as shown in Tables 2 and 3.

Table 2

LIQUID PROCESSING
Tertiary Sequence

1. Primary treatment or gross solids separation
2. Secondary treatment or biological oxidation
3. Chemical treatment and phosphorus removal
4. Nitrogen removal by ammonia stripping
5. Mixed media filtration
6. Granular activated carbon adsorption of dissolved organics
7. Disinfection by chlorine

Table 3

SOLIDS HANDLING

1. Incineration of waste primary, secondary, and chemical sludges.
2. Regeneration and reuse of spent granular carbon.
3. Recalcination and reuse of lime mud.

This tertiary sequence of treatment is designed to provide an exceptionally high degree of treatment. Typical quality of the water is shown in Table 4.

Table 4

TYPICAL QUALITY OF RECLAIMED WATER

Description	Average Value
B.O.D.	less than 1.0
C.O.D.	10
Suspended Solids mg/l	0
Turbidity, J.U.	0.3
M.B.A.S., mg/l	0.10
Phosphorus, mg/l as P	0.10
Color, Units	<2
Odor	None
Coliform Bacteria M.P.N./100 ml.	<2.0

The results of treatment by this method are very uniform, and operation of the plant is extremely reliable. The plant has operated continuously and without interruption for 2-1.2 years, and the water quality has always exceeded the high standards established by the regulatory agencies. The reliability corresponds to that obtained in power plants and water purifications plants in contrast to the lower level common in most secondary sewage treatment plants. The tertiary treatment makes up for upsets and shortcomings of the secondary treatment.

This plant scale demonstration of the reliability of these new treatment processes satisfactorily answers questions many people had regarding the practical operation of what might be considered a rather complex process. This accomplishment is the result of two basic essentials: (1) adequate financing to do the job, and (2) thorough on-the-job training of all plant operators prior to plant startup. Our experience clearly shows the extreme importance of proper attention to these facets of an advanced wastewater treatment project.

Our operator training program consisted of 2 hour sessions, 5 days per week for a period of 3 months. Consulting engineers, equipment suppliers, manufacturer's representatives, and construction contractors all participated in supplying information pertinent to plant operation. A complete 200 page training manual which was prepared by the Consulting Engineer specifically for the Tahoe plant was the basic reference for the training course. The information in the manual was supplemented by literature and data from suppliers and manufacturers. An adequate operations manual and on-the-job training for plant personnel are necessities.

At Indian Creek Reservoir, which receives the reclaimed water, a Rainbow Trout fishery has been established which is one of the best in the area.

The actual construction and operating costs for the Tahoe project are shown in Table 5. Costs for the high degree of advanced treatment are about the same as for conventional secondary treatment.

The Tahoe Process can be scaled up or down to virtually any desired plant capacity. For larger or smaller plants, there would be differences in types of structures and equipment for best economy. At capacities less than 2 mgd, powdered carbon would be substituted for granular carbon and lime recalcination would not be practical.

To meet less stringent quality requirements, there are an almost unlimited number of variations which could be made to meet the applicable standards.

Acknowledgements. Much of the text contained in this paper is taken from a textbook and design manual on "Advanced Wastewater Treatment" by Russell L. Culp and Gordon L. Culp, which is now in publication by Van Nostrand Reinhold Co., and which will be released early next year.

The Tahoe plant is a Demonstration Project of the Federal Water Quality Administration, who rendered financial assistance in the construction and data collection program.

Table 5

SUMMARY OF
TOTAL COSTS EXPERIENCED
AT SOUTH LAKE TAHOE

Process	Total Operating Costs	Capital Costs	Total Costs $/mg Influent
Primary	11	19.50	30.50
Activated Sludge	31	33.60	69.60
Organic Sludge	65	14.10	79.10
Chlorination	2	0.30	2.30
Lime Treatment	44	9.70	53.70
Lime Recalcining	32	13.50	45.50
Ammonia Stripping	11	8.00	19.00
Recarbonation	2	4.00	6.00
Filtration	22	17.80	39.80
Carbon Adsorption	3	16.30	19.30
Carbon Regeneration	13	5.20	18.20
Total of Conventional Treatment	109	67.50	176.50
Total of Advanced Wastewater Treatment	127	74.50	201.50
TOTAL	$ 236	$142.00	$ 378.00

Nitrogen Removal from Agricultural Wastewater

by

Louis A. Dean

INTRODUCTION

In recent years an ever increasing emphasis has been placed on nitrogen removal techniques, both for reasons of public health and the control of eutrophication. The subject studies were instituted because of the possible eutrophying effect of the discharge of large quantities of agricultural tile drainage into the San Francisco Bay System. The tile drainage to be disposed of results from the irrigation of large portions of California's San Joaquin Valley that have been classified as drainage problem areas (Figure 1). Of the over 7 million acres (2.8 million ha) of irrigable land in the Valley, approximately 1.7 million acres (0.7 million ha) have been classified as existing or potential drainage problem areas -- lands where brackish high water tables have or will become a problem (1).

Most of the saline agricultural wastewater that is presently being generated within the Valley is discharged to the San Joaquin River. As the quantity of this waste increases, due to the increased availability of water for irrigation, the waste load on the river would reach a point where downstream users would be deprived of a usable water supply.

For this reason, a system is now under construction to collect much of the saline drainage and transport it from the Valley. The system will serve 300,000 acres (120,000 ha) of the drainage problem area. It is being built by

---

Louis A. Dean is Senior Sanitary Engineer, California Department of Water Resources, Sacramento, California.

Reproduced from best available copy.

the U. S. Bureau of Reclamation (USBR) to handle primarily the wastewater from the Federal San Luis Service Area. California's Department of Water Resources (DWR) is also aware of the drainage problem and is making plans to serve those areas outside of the Federal San Luis Service Area. The predicted quality of the waste to be carried by the San Luis Drain is presented in Table 1 (2). It is estimated that by the year 2000 there will be 500,000 acre-feet (6.2×10^8 cu m) of tile drainage to be disposed of annually. (This is only about 4 percent of the water to be applied to the Valley in the same period.) The peak flows, during the summer irrigation season, are expected to reach 700 mgd (2.65×10^6 cu m/day).

Prior to the study reported herein, the DWR and USBR had investigated disposal of tile drainage for several years. Of all the alternatives studied, they concluded that the most economical and safest method of disposal would be a gravity concrete-lined canal, or drain, that would discharge into the upper reaches of San Francisco Bay (3). They further concluded that the only major constituent of the drainage that could create problems in the receiving water is nitrogen. A 1967 Federal Water Quality Administration (FWQA) study of water quality problems which could result from the drainage water concurred with the earlier study (4). The FWQA concluded that problems would not occur if the nitrogen content of the subject waste could be reduced to 2 mg/l or less.

The DWR, USBR, and FWQA joined together to determine if a method of nitrogen removal for tile drainage could be developed or if the costly alternative of ocean disposal would have to be adopted. In order to evaluate possible methods of nitrogen removal under field conditions, the Interagency Agricultural Waste Water Treatment Center was established. It is located near Firebaugh, California (40 miles west of Fresno) in one of the existing drainage problem areas. This paper presents a comparison of the various treatment methods under study at the Center.

TREATMENT TECHNIQUES

Ninety-five percent, or more, of the nitrogen in tile drainage is in the nitrate form. Therefore, the problem of reducing the nitrogen content of tile drainage becomes one of removing nitrate-nitrogen. Two basic methods of nitrate removal are being investigated: bacterial denitrification, and algal biomass production and harvesting (algae stripping). Bacterial denitrification is being studied in both anaerobic filters and deep ponds.

Bacterial Denitrification

Bacterial denitrification is a process in which bacteria utilize highly oxidized anions for the oxidation of organic matter (5). The overall process can be expressed stoichiometrically, as a two-step or three-step reaction. Nitrate (NO_3^-) is reduced to nitrite (NO_2^-) and then from nitrite to molecular nitrogen (N_2) with or without the intermediate formation of hyponitrite (N_2O_2). The process can be shown to be thermodynamically sound (capable of producing energy for organisms utilizing it) (6). Several commonly occurring facultative bacteria are denitrifiers, for example, Bacillus brevis[1/] The occurrence of bacterial denitrification in existing sewage treatment plants is well documented and several pilot-scale units have been built and successfully operated for denitrification (7, 8, 9, 10).

Unless biodegradable organic material is present in adequate quantity, effective denitrification will not take place. Most agricultural tile drainage has a low organic content. Therefore, methanol is added to the waste prior to entry into the units operated at the Center (11).

The main factors, with regard to bacterial denitrification, requiring study were: the general feasibility of the process, detention time, methanol requirement, and operating efficiencies under field conditions.

1/ One of the three bacteria of the Bacillus genus which are cultured commercially to produce polypeptide-type antibiotics.

Anaerobic Filters

In the anaerobic filter the waste is passed through a packed column-type reactor (a vessel containing an inactive medium). The filter medium provides a solid support for the denitrifying bacteria. The advantage of the filter is that the solids retention time of the unit is longer than the hydraulic detention time and, therefore, there is no need for the recycling of solids (12). Anaerobic filters can be operated with upflow or downflow configurations, as long as the unit is kept anaerobic (or nearly so). All of the filters operated at the Center have been upflow units (Figure 2).

Four different sizes of anaerobic filters have been used: 4-, 18-, and 36-inch (10-, 46-, and 91-cm) diameter (13) and 10-feet by 10-feet (3-m X 3-m) square. The 10-feet by 10-feet filter has a false bottom (of the type used in water treatment rapid sand filters) with an 8-inch (20-cm) plenum and a 6-foot (1.8-m) bed depth. The primary purpose for building this filter was to investigate the effect of scale-up on process efficiency due to changes in the hydraulic regime.

Reproduced from best available copy.

Deep Ponds

The feasibility of bacterial denitrification in deep ponds under field conditions was determined in 3-foot (0.9-m) diameter simulated deep ponds (14). Once the process was shown to be feasible, two large ponds were constructed as shown schematically in Figure 3. The larger of the two ponds constructed at the Center is 200-X 50-feet (61-X 15-m) and has a floating cover. The smaller pond is 50-X 50-feet (15-X 15-m) and is not covered. Both ponds are approximately 14-feet deep (4.3-m). Through the parallel operation of these two units, it was possible to evaluate the significance of wind mixing, and algal growth on process efficiency. The solids retention time and hydraulic detention time are the same, therefore, the controlling parameter in pond operation is the regeneration time of the bacterial population. To increase the solids retention time, recycle is

necessary. If an adequate residence time is not provided, effective denitrification is not maintained.

Algae Stripping

Research on the treatment of domestic and industrial wastes in photosynthetic systems has been studied for a number of years (15). As a result of the findings of such studies, it has become apparent that nutrients are assimilated by algal cells in significant amounts and, hence, are removed as part of the operation.

Figure 4 is a schematic diagram of algae stripping. It is immediately evident, upon observing any ponded tile drainage, that algae definitely grow in this type of waste. The main problem in the process is not the mere growing of algae, but the optimization of growth factors to maximize their nitrate uptake rate. Some of the factors that must be investigated before an algal growth system can be evaluated are: growth pond depth, hydraulic detention time, recirculation, mixing, pH, and nutrient addition. All of these factors must be optimized before a photosynthetic system can be designed. Once the algae have been grown, the problem of separating them from the waste becomes paramount. Flocculation, sedimentation, flotation, and microscreening are all processes that should be evaluated to determine the least costly technique.

Three types of growth vessels are used at the Center: a light box, small-scale growth ponds (miniponds), and a 1/4-acre (0.10-ha) large-scale rapid growth pond (16). The light box is used for the preliminary screening of factors such as pH and nutrient addition. These results are then verified in the miniponds. In addition to varifying light box results, miniponds are used to study detention times, mixing, water depth, and climatic conditions. Final testing is conducted in the large-scale rapid growth pond. The pilot-scale growth pond also serves as the primary source of algal-laden waste for the separation studies.

The harvesting studies were divided into three categories--separation, dewatering, and drying. The majority of the work that has been done on separation has consisted of jar tests to evaluate various flocculants. Pilot-scale separation techniques studied included: centrifugation, microscreening, filtration, and flocculation and sedimentation. The following dewatering and drying equipment were investigated: cylinderical and basket-type centrifuges, a vacuum filter, and sand drying beds.

Analytical Techniques

The routine analyses used to monitor the operation of the various units at the Center are summarized as follows:

ROUTINE ANALYSES PERFORMED AT THE
INTERAGENCY AGRICULTURAL WASTE WATER TREATMENT CENTER

Analyte	Technique
Nitrate-Nitrogen	Brucine Method and/or Selective Ion Electrode
Nitrite-Nitrogen	Standard Methods, 12th Edition (17)
Total Kjeldahl Nitrogen	Kjeldahl Method
Ammonia Nitrogen	Distillation Method
Organic Nitrogen	Kjeldahl Method
Orthophosphate	Stannous Chloride Modification
pH	Glass Electrode
Alkalinity	Standard Methods, 12th Edition
Dissolved Oxygen	Winkler Method
Suspended Solids	0.45 μ Glass Paper, 103°C
Volatile Suspended Solids	0.45 μ Glass Paper, 600°C
Optical Density	450 μ, 5 cm Cell
Electrical Conductivity	Galvanic Cell
Algal Cell Counts and Identifiers	Sedgwick-Rafter Cell
Methanol	Gas Chromatograph, Carbowax Column, Flame Ionization Detector

Most samples were normally collected between 8:00 and 9:00 AM and analyzed immediately. Some samples were taken in the afternoon to gather information on changes that may occur during the peak photosynthetic period. Diurnal studies were also conducted, as the need arose.

In addition to chemical analyses, several physical parameters were monitored. Water temperatures were monitored daily with maximum-minimum thermometers

and periodically with 8-day recording thermographs. The influent pressure required for an anaerobic filter to maintain a constant hydraulic detention time was monitored with varying frequency throughout the study. Flows were calibrated volumetrically. In addition tracer studies using the chloride ion as a tracer were run to determine the actual hydraulic regime of the different units. Pre- and post-injection density corrections were made using sodium sulfate. The tracer studies were analyzed using the volume apportionment technique (18, 19).

Influent Quality

The tile drainage used for the investigation reported in this paper came from a drainage system that was in existence when the Center was constructed. This system, which serves approximately 400-acres (160-ha), was chosen because of its proximity to land available for construction of the Center and the quality of the drainage, which was considered to be quite "typical". The range of constituent concentrations in this drainage are as follows:

CHARACTERISTICS OF TILE DRAINAGE USED
AT THE INTERAGENCY WASTE WATER TREATMENT CENTER

Constituents	Range of Concentrations mg/l
Total Dissolved Solids	2500-7600
Salts	
Sulfate	1500-3900
Sodium	620-2050
Chloride	310-640
Calcium	160-390
Magnesium	70-230
Bicarbonate	280-330
Boron	4-15
Potassium	4-11
Nutrients	
Nitrogen	5-25
Phosphate	0.13-0.33
Pesticides	0.001
Others	
5 Day BOD	1-3
COD	10-20
DO	7-9

The variations are seasonal and due primarily to irrigation applications and rainfall. With the exception of the variation in nitrogen concentration, it was decided that all other constituents should be allowed to vary naturally. Sodium nitrate was used to maintain an average influent nitrate-nitrogen concentration of approximately 20 mg/l, the predicted average nitrogen concentration of the waste in the San Luis Drain.

DISCUSSION

Figure 5 illustrates the predicted seasonal variations of flow and nitrogen concentration. The distribution of flow was determined through a monitoring program and computer similation (3). The nitrate-time relationship was developed from data collected by the DWR in a special three year water quality monitoring program (20). Figure 6 is a summary of westside San Joaquin Valley air temperature data from three stations in the general area in which a treatment plant might be constructed (between Los Banos and Tracy, California) (21). These figures are presented to illustrate the fact that the majority of the waste load will occur during the warmer summer months. Seventy percent of the annual nitrogen load, in pounds per day (grams/day), will arrive at the plant between April 1 and September 30. The significance of this lies in the fact that all of the nitrogen removal systems being evaluated at the Center are biological systems, and are temperature sensitive.

Operational data will be presented for each system to illustrate individual operating criteria, etc. The individual systems will then be compared.

Anaerobic Filter

The data from experiments designed to investigate the significant media size, texture, and sorptive quality are summarized below (13). Based on the results of an initial feasibility investigation, all filters operated to generate this data had hydraulic detention times (based on void volumes) of 2 hours. Medium surface texture, size, and sorptive quality have no apparent affect on removal

efficiencies. After an extended period of continuous operation the backwash rate within filters containing media with diameters of less than 1-inch (2.54-cm) built up to the point where the required influent pressures (as high as 60-psig) rendered them uneconomical.

NITROGEN REMOVAL EFFICIENCIES FOR FILTER DENITRIFICATION UNITS CONTAINING VARIOUS MEDIA

Medium	Nitrogen Removal Efficiency, Percent		
	Min.	Max.	Average
Activated Carbon	89	99	96
Washed Sand	84	97	93
5/16" Coal	80	98	93
5/16" Volcanic Cinders	85	98	94
3/8" Aggregate	82	97	94
5/8" Volcanic Cinders	87	97	91
1" Coal	81	98	93
1" Volcanic Cinders	89	97	96
1" Aggregate	89	98	94

The data gathered from the long-term operation of three filters containing 1-inch diameter media are summarized in the following table.

SUMMARY OF LONG-TERM PERFORMANCE OF FILTERS CONTAINING 1-INCH DIAMETER AGGREGATE

Detention Time Hrs.	Days of Operation	Percent Nitrogen Removal			Required Influent Pressure, psig		
		Min.	Max.	Average	Min.	Max.	Average
0.5	275	40	91	68	3.5	11.8	7.2
1.0	240	64	97	88	3.2	9.6	5.4
2.0	244	79	97	91	3.5	9.7	6.2

For about the first 150 days of operation, the required influent pressure fluctuated within a narrow range. There did not appear to be any need for backwashing, etc. Since that time, an upward trend has begun to appear. Investigations are currently underway to determine the best technique to reduce the pressure.

The temperature of the influent to the units varied from a high of 22° to a low of 10° Centigrade. There were some indications of a relationship between temperature and nitrogen removal efficiency; however, the relationship was

far from obvious. By grouping the nitrate and nitrite removal data for samples taken at different depths within a filter, the reduction in efficiency becomes more apparent as illustrated in the next table.

PROFILE DATA GROUPED BY PERCENT OF FILTER REQUIRED FOR 80% OR GREATER NO_3+NO_2 REDUCTION

(1-inch diameter aggregate media, 2-hour hydraulic detention time)

Percent of Filter Used	Average Influent Temperature °C
25% or less	14.5
25% to 50%	12.8
50% to 75%	11.9

As the temperature of the influent dropped, the percent of the filter bed required for the same degree of treatment increased. The 6-foot bed depths used for this investigation and the range of temperatures encountered were such that the effect was not reflected in the overall treatment efficiencies of the units. This information is only qualitative, since steady-state conditions were not reached at any given temperature. Additional study is currently (October 1969) being undertaken to quantify the effect of temperature on filter denitrification. This study is being conducted using a filter which was constructed in a low temperature incubator.

The following equation was developed by McCarty to express the concentration of methanol required for denitrification (11).

$$C_m = C_r (1.90 N_o + 1.177 N_1 + 0.67 D_o) \qquad (1)$$

C_m = required methanol concentration, mg/l

C_r = consumptive ratio (determined experimentally)

C_r = total quantity of organic carbon source consumed during denitrification divided by the stoichiometric requirement for denitrification and deoxygenation

N_o = initial nitrate-nitrogen concentration, mg/l

N_1 = initial nitrite-nitrogen concentration, mg/l

D_o = initial dissolved oxygen concentration, mg/l

herent difficulties in [illegible]. With the development of [illegible] which would respond to changes in influent [illegible] could be drastically reduced. The [illegible] nitrate-nitrogen and a [illegible] by equation (1) equals 64 mg/l.

Deep Pond

The initial [illegible] were conducted in the [illegible] ponds were operated at various detention [illegible] denitrification took place in ponds with [illegible].

Data gathered [illegible] of 1968, are summarized in Table [illegible] consistently outperformed uncovered ponds. There are [illegible] develop in uncovered ponds [illegible] oxygen concentration inhibits denitrification. [illegible] is short-circuited through the unit [illegible] temperature variable in an uncovered pond. The high [illegible] provided for the uncovered pond when operated with a [illegible] primarily caused by the formation of a natural partial cover. [illegible] of entrapped algal and bacterial cells floating on the [illegible] type of covering is necessary for sufficient [illegible] currently underway to determine if a [illegible] in place of fabricated pond covers.

[illegible]

with these units; therefore, it is, as yet, impossible to accurately predict the minimum detention time possible. However, a ten-day detention time, with 25% recycle, and a pond depth of 14-feet (4.27-m) have been shown to be effective for covered pond summertime operation (Table 2) and are used in this paper.

The methanol requirement for pond denitrification follows the same equation developed for filter denitrification. The average consumptive ratio for deep ponds has been found to be statistically equal to that found for anaerobic filters, within the accuracy possible under field conditions.

Algae Stripping

The results of the algae stripping studies are divided into three subsections; growth, harvesting, and algae utilization.

Growth. Light box studies conducted early in the investigation indicated a serious phosphorus limitation in tile drainage. The addition of approximately 2.0-mg/l of phosphorus was found necessary. This was further confirmed in studies in the miniponds.

Once the required phosphorus addition was determined, work began on optimizing nitrate assimilation. Of all the nutrient additions investigated to date, iron and carbon appear to be the only two that have significant effects on assimilation. The optimum amount of iron that should be used has been found to be approximately 2.5 mg/l. A possible seasonal variability in the iron requirement has been indicated and 2.5 mg/l appears to be a reasonable maximum requirement. By using a mixture of carbon dioxide (4%) and air in light box studies, nitrate assimilation has been improved. Carbon dioxide studies in the miniponds have been initiated, however, because of mechanical problems the results are, as yet, only qualitative.

Physical parameters which have been studied that effect the process are: water depth, mixing, and hydraulic detention time. Water depths of from 8- to 16-inches (20.3- to 40.6-cm) have been studied in combination with various deten-

several techniques operated to study the significance of these parameters. This table is intended to show the effects of physical parameters only. The low efficiencies resulted because the other factors that affect growth were not optimized. For continuous operation a design depth of 12-inches (30.5-cm) with 4 1/2-hours of mixing and a detention time of 5-days has been found to be adequate. Because of the difference between summer and winter temperatures and light a 10- to 15-day detention time is required in the winter.

Harvesting. Oswald, et. al., determined that flocculation and sedimentation was the most economical and efficient method of algae separation at the time of their study (22). Consequently, the first work done on separation at the Center involved jar tests. The following flocculants were investigated: alum, lime, ferric chloride, ferric sulfate, and various polyelectrolytes. Better than ninety per cent removal of algae cells from the liquid has been achieved with alum, lime, ferric sulfate, and several polyelectrolytes. The most economical flocculant used thus far has been ferric sulfate.

Studies utilizing pilot-scale separation equipment indicated that greate than 95 percent of algae cells is possible. These units have been operating for a short period of time. The economics and efficiencies of the pilot-scale harvestin units have not been determined at this time.

Algae Utilization. At peak flows more than one million pounds (4.5×10^5-kg) of dry algae will be produced each day. The use of algae as a protein supplement in the feed formulations for sheep and hogs has been shown to be feasible (22). A study investigating the use of algae for poultry feed determined that in addition to the value of the protein content, a benefit is derived from the presence of significant quantities of Xanthophyll (22). (A carotenoid which is used to increase the color of egg yolks and bird flesh to enhance market value. Other possible uses for algae include; soil conditioning or fertilization, and

production of glues or adhesives. Depending on use, algae have been found to have a value of between 80 and 160 dollars per ton (912 kg).

COMPARISON OF METHODS

The pilot-scale studies at Pirebaugh have demonstrated that all three methods can remove ninety percent of the nitrogen (mostly nitrate) in agricultural wastewaters. Differences in the biological systems and physical configurations of the processes result in many differences between the methods, Table 4.

Anaerobic filters operate efficiently with the shortest detention times of the three methods studied, one hour, because of the concentrating and containing of bacteria within the units. Algae are not as readily washed out of ponds as bacteria. Therefore, algae growth ponds operate at a five-day detention time compared to ten-days for deep pond denitrification. Anaerobic filters, with media depths of six-feet and detention times of one-hour, require 0.053-acres (0.021-ha) or 2300-square feet (214-square m) of surface area for each million gallons per day (3,785-cu m/day) of waste treated. The deep pond denitrification process requires 2.0-acres (0.81-ha) of ponds to treat the same quantity of waste. This requirement was calculated using a pond depth of 14-feet (4.3-m) and detention time of 10-days. The growth pond surface area needed for algae stripping has been determined to be 15.3-acres (6.2-ha)with a one-foot (0.3-m) water depth and 5-day detention time. For a design flow of 700 million gallons per day, the relative surface areas would be 37-acres (15-ha) for filters, 1400-acres (570-ha) for denitrification ponds and 10,700-acres (4,300-ha) for algal growth ponds. All three treatment methods require some chemical additions. The denitrification systems will require 82-gallons (310-l) of methanol for each million gallons per day of agricultural tile drainage. For each million gallons per day the algal system will require 17-pounds (7.7-kg) of phosphorus, 21-pounds (9.5 kg) of iron, and 250-pounds (113-kg) of carbon dioxide for growth. The separation portion of the algae stripping process will require 41-pounds (18.6-kg) of ferric sulphate per million gallons per day. Algae

aeration equipment for bacterial denitrification. Chemical feed pumps for chemical addition, influent pumps for anaerobic filters, and recycle pumps for deep ponds are the major pieces of mechanical equipment required for bacterial denitrification. Because of the low dissolved oxygen content of wastes that have been treated by these processes, it may be necessary to reaerate the effluent prior to discharge. Algal stripping will require phosphorus, iron, and carbon dioxide feed systems; growth pond mixing systems; separation equipment; dewatering equipment; drying equipment; and equipment to handle the algae to prepare it for marketing.

Each of the methods has special operational requirements. The solids level in denitrification ponds and algal growth ponds must be maintained at a level high enough to achieve the necessary nitrogen removal efficiency. The solids level in the filters must be maintained low enough to prevent clogging and excessive head loss through the units. The chemical feed rates have to be carefully monitored to provide the required additive concentrations for the biological cultures, as economically as possible. The three methods are temperature sensitive, and as temperatures vary throughout the year, detention times will have to be adjusted. The separation procedures may vary throughout the year as the nature of the algal culture changes.

Algae stripping is the only method that produces a by-product. Approximately one ton of dry algae will be produced for each million gallons per day of waste treated. The value of the algae produced will be between 80 and 160 dollars per ton.

The preliminary feasibility reports for bacterial denitrification and algae stripping estimated costs to be between 25- and 30-dollars per million gallons for each treatment method (11, 24). Preliminary cost estimates based on the feasibility studies will be in 70- and 110-dollars per million gallons for each treatment method.

In conclusion, all of the methods have several advantages and disadvantages. The most desirable system will probably be a combination of algae stripping and bacterial denitrification. In a combined process, algae growth would be encouraged in the drain, and a combination discharge control reservoir and growth pond. The algae would then be separated from the waste and processed for marketing. A bacterial denitrification unit would provide final polishing to insure that the nitrogen level has been reduced to that required for discharge to the San Francisco Bay System. Such a plan would have the advantage of a dual system. The required nitrogen removal would always be accomplished. It would also have the advantage of a by-product income to affect treatment costs.

References

1. Stetson, C. L., "A Drainage System for the San Joaquin Valley," 4th International Water Quality Symposium, San Francisco, California (August 14, 1968).
2. Price, E. P., "The San Luis Drain," 4th International Water Quality Symposium, San Francisco, California (August 14, 1968).
3. Lindholm, R. R., "San Joaquin Valley Drainage Investigation-San Joaquin Master Drain", California Department of Water Resources, Bulletin No. 127, Preliminary Edition (January 1965).
4. "Effects of the San Joaquin Master Drain on Water Quality of the San Francisco Bay and Delta," U.S. Department of the Interior, Federal Water Pollution Control Administration, Southwest Region, San Francisco, California (January 1967).
5. Hutchinson, G. E., A Treatise on Limnology-Volume 1, John Wiley and Sons, Inc., New York, New York (1957).
6. McCarty, P. L., Beck, L. A., and St. Amant, P. P., "Biological Denitrification of Waste Waters by Addition of Organic Materials," 24th Industrial Waste Conferen Purdue University, Lafayette, Indiana (May 1969).
7. Christenson, C. W., et al., "Reduction of Nitrate-Nitrogen by Modified Activated Sludge," USAEC, TID. 7517, Pages 264-267 (1956).
8. Ludzack, F. J., and Ettinger, M. B., "Controlling Operation to Minimize Activated Sludge Effluent Nitrogen," Journal Water Pollution Control Federation, Vol. 34, No. 9, Pages 920-931 (1962).
9. Bishop, D. F., "Ammonia and Nitrogen Removal," Internal Report, Advanced Treatment Research, FWPCA, R.A. Taft Water Research Center, Cincinnati, Ohio.
10. Finsen, P. O., and Sampson, D., "Denitrification of Sewage Effluents," Water and Waste Treatment Journal, Vol. 7, pages 298-300 (1959).

11. McCarty, P. L., "Feasibility of the Denitrification Process for Removal of Nitrate-Nitrogen from Agricultural Drainage Waters," Appendix California Department of Water Resources Bulletin 174-3 (May 1969).

12. Young, J. C., and McCarty, P. L., "The Anaerobic Filter for Waste Treatment," Stanford University, Technical Report No. 87, Stanford, California (March 1968).

13. Tamblyn, T. A., and Sword, B. R., "The Anaerobic Filter for the Denitrification of Agricultural Subsurface Drainage," 24th Industrial Waste Conference, Purdue University, Lafayette, Indiana (May 1969).

14. Brown, R. L., "Field Evaluation of Anaerobic Denitrification in Simulated Deep Ponds," California Department of Water Resources Bulletin No. 174-3 (May 1969).

15. Golueke, C. G., Oswald, W. J., and Gee, H. K., "Increasing High-Rate Pond Loading by Phase Isolation-Final Report," Sanitary Engineering Research Laboratory, University of California, Berkeley, California (April 1962).

16. Beck, L. A., Oswald, W. J., and Goldman, J. C., "Nitrate Removal from Agricultural Tile Drainage by Photosynthetic Systems," Second National Symposium, Sanitary Engineering Research Development and Design, American Society of Civil Engineers, Cornell University, Ithaca, New York (July 1969).

17. "Standard Methods for the Examination of Water and Wastewater", American Public Health Association, Inc., New York, New York, 12th Edition (1965).

18. Cholette, A., Blanchet, J., and Cloutier, L., "Performance of Flow Reactors at Various Levels of Mixing," Canadian Journal of Chemical Engineering, Vol. 38, (1960).

19. Milbury, W. F., "A Development and Evaluation of a Theoretical Model Describing the Effects of Hydraulic Regime in Continuous Microbial Systems," dissertation presented to Northwestern University at Evanston, Illinois, in 1964, in partial fulfillment of the requirement for the degree of Doctor of Philosophy.

20. California Department of Water Resources, San Joaquin District, Fresno, California, unpublished data.

21. California Department of Water Resources, "Hydrologic Data, Volume IV: San Joaquin Valley, Appendix A, Climatological Data," Bulletin 130, (September 1968).

22. Golueke, C. G., and Oswald, W. J., "Harvesting and Processing of Sewage-Grown Planktonic Algae," Journal Water Pollution Control Federation, Vol. 37, No. 4, Pages 471-498 (1965).

23. Anonymous, "A Study of the Use of Biomass Systems in Water Renovation", North American Aviation Inc., (May 1, 1967).

24. Oswald, W. J., Crosby, D. G., Golueke, C. G., "Removal of Pesticides and Algal Growth Potential from San Joaquin Valley Drainage Waters (A Feasibility Study), California Department of Water Resources unpublished report (1964).

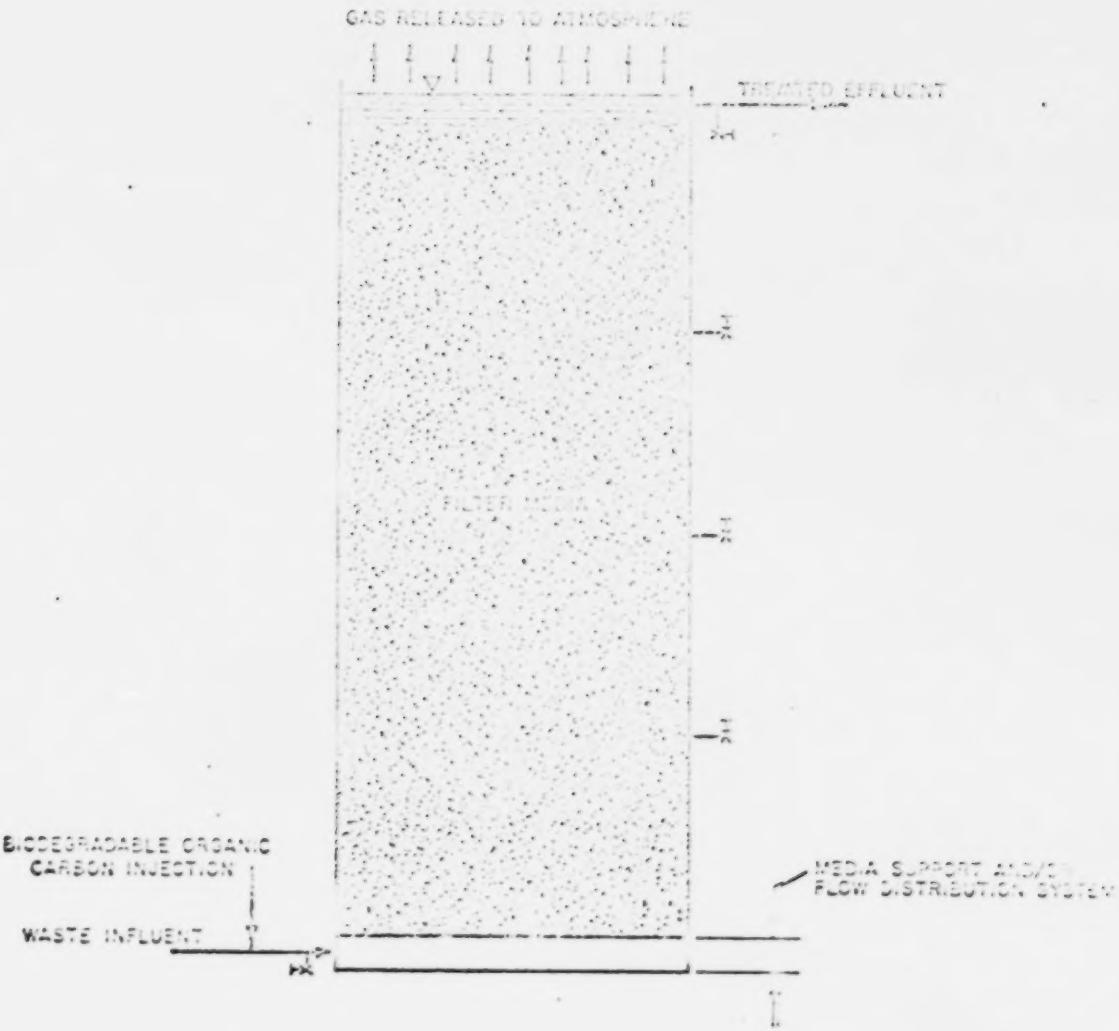
EXISTING AND POTENTIAL AGRICULTURAL WASTE WATER DISPOSAL PROBLEM AREAS
SAN JOAQUIN VALLEY- CALIFORNIA

SCALE OF MILES
40 0 40 80 120 160 200

TA[illegible] - [illegible] Concentrations of

Chemical Substance in Raw Main Drain Waters

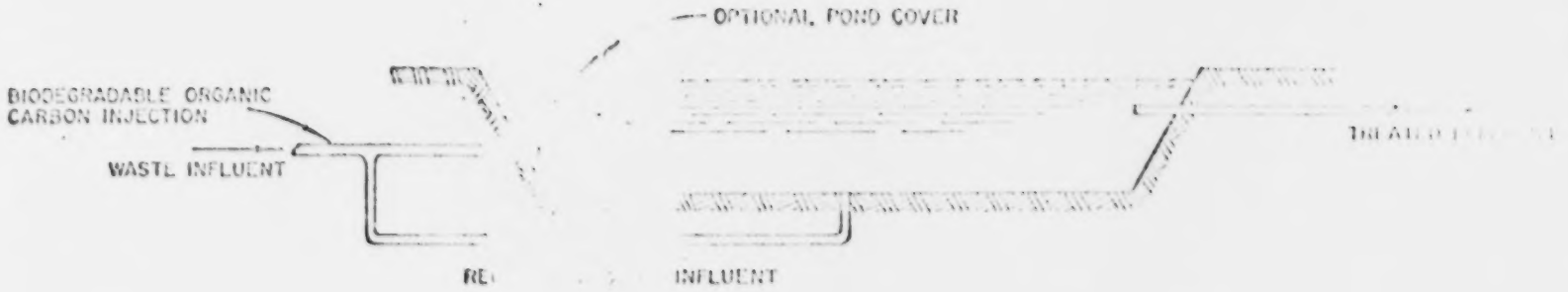
Substance	1970 Conditions mg/l	2000 Conditions mg/l
Total Dissolved Solids	7,000	3,000
Salts		
Sulfate	3,000	700
Sodium	1,500	700
Chloride	1,200	900
Calcium	300	100
Magnesium	200	50
Bicarbonate	200	100
Potassium	20	10
Boron	10	3
Nutrients		
Total Nitrogen	20	20
Total Phosphate	0.15	0.15
Pesticides	< 0.001	< 0.001
Others		
5 Day B.O.D.	1-3	1-3
C.O.D.	10-20	10-20



LEGEND

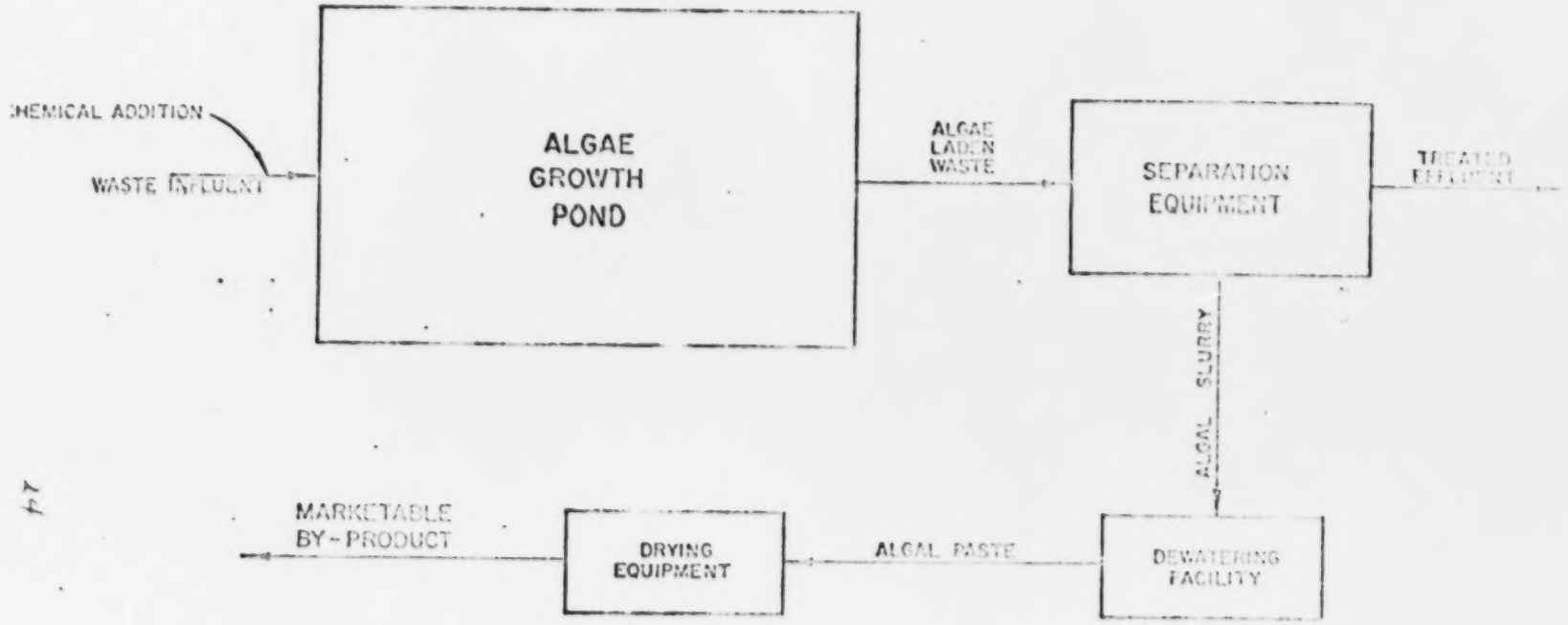
OPTIONAL SAMPLE TAPS

SCHEMATIC DIAGRAM
UPFLOW ANAEROBIC FILTER DENITRIFICATION PROCESS

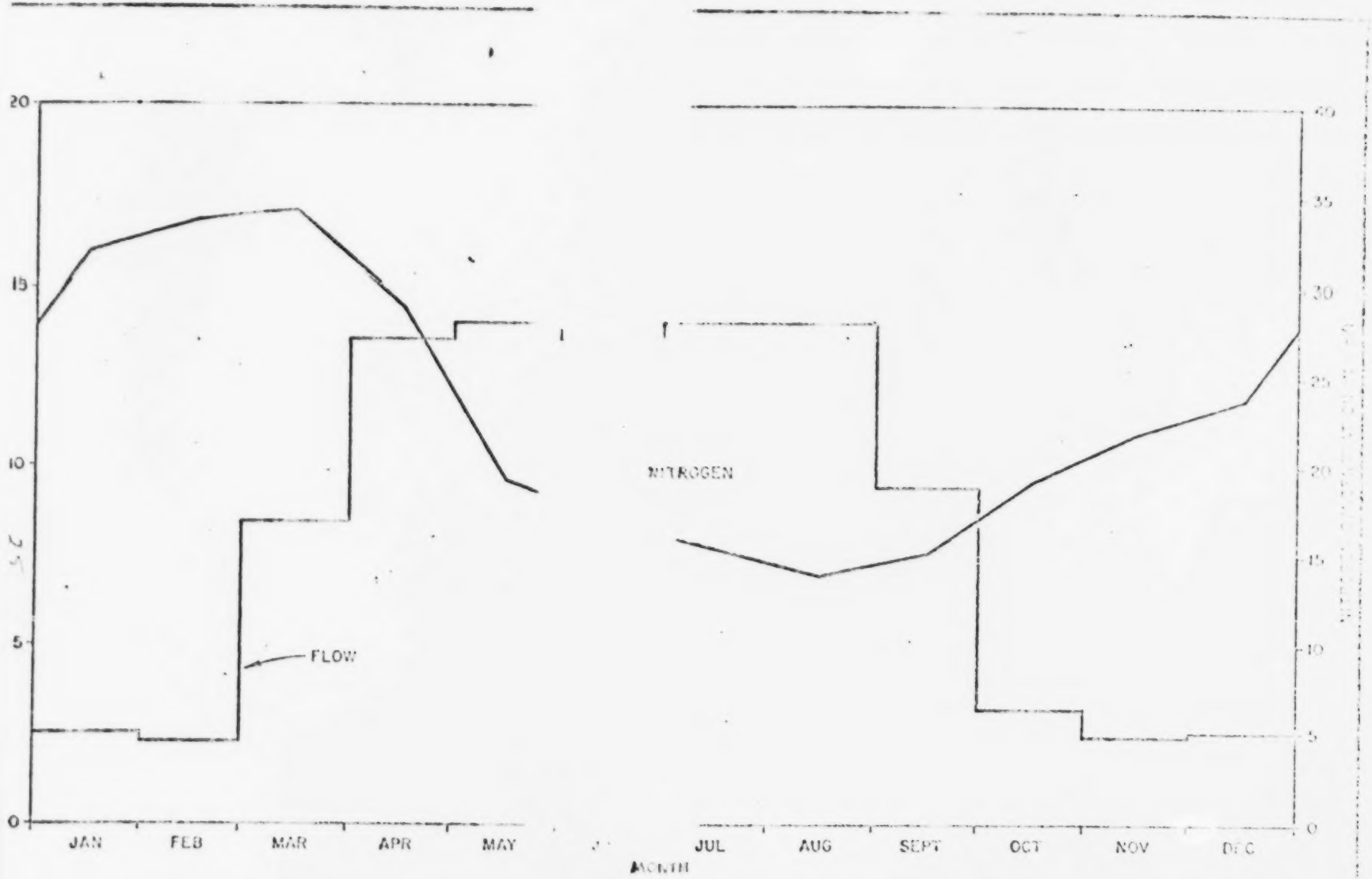


Reproduced from best available copy.

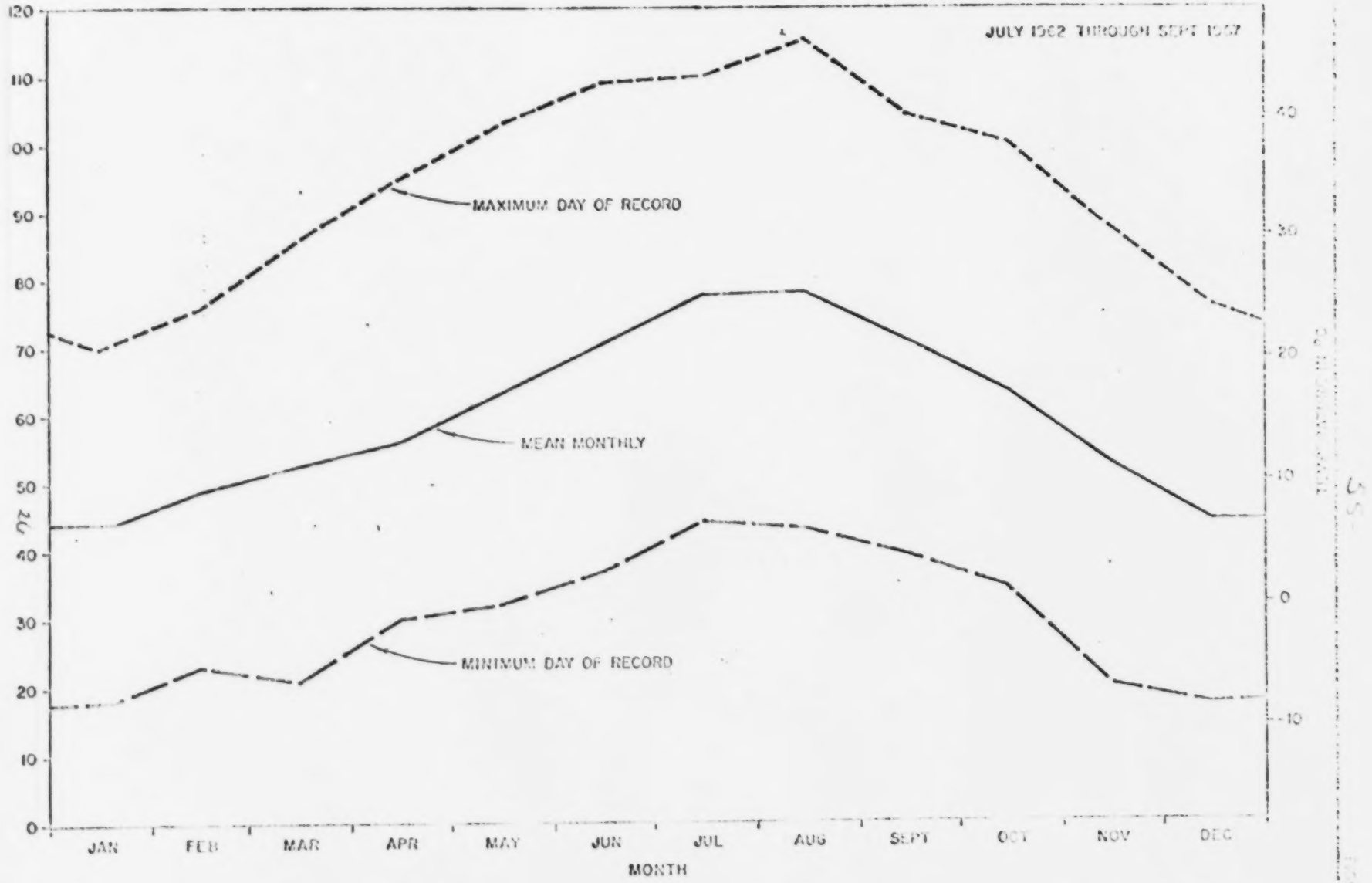
SCHEMATIC DIAGRAM
DENITRIFICATION PROCESS



SCHEMATIC DIAGRAM
ALGAE STRIPPING PROCESS



PREDICTED SEASONAL VARIATION OF THE DRAINAGE FLOW & NITROGEN CONCENTRATION FROM SAN JOAQUIN VALLEY, CALIFORNIA



ANNUAL AIR TEMPERATURE VARIATION ON THE WEST SIDE OF THE SAN JOAQUIN VALLEY, CALIFORNIA

TABLE 2

Nitrogen Removal-Detention Relationship for Deep Pond Denitrification

Detention Time Days	Number of Days of Operation		Nitrogen Removal in Percent					
			Uncovered Pond			Covered Pond		
	Uncovered Pond	Covered Pond	Min.	Max.	Avg.	Min.	Max.	Avg.
7.5	-	34	-	-	-	67	92	
10	61	31	60	93	77	75	93	
15	30	35	30	95	56	88	94	
20	31	40	54	84	71	84	94	

TABLE 3

Nitrate Removal Efficiency of Algae Stripping Growth Units as Related to Physical Parameters

Detention Time				Mixing				Depth			
Days	Nitrate Removal in %			Hours Per Day	Nitrate Removal in %				Nitrate Removal in %		
	Max.	Min.	Average		Max.	Min.	Average	Inches	Max.	Min.	Average
2.5	75	43	59	0	-	-	59	8	98	71	83
5.0	80	59	75	4.5	87	78	83	12	87	74	79
8.0	82	74	78	14	85	74	79	16	87	61	75
10.0	90	80	85	24	-	-	89				

MINERAL REMOVAL BY ION EXCHANGE, REVERSE OSMOSIS AND ELECTRODIALYSIS

Franklin D. Dryden
Deputy Assistant Chief Engineer
Sanitation Districts of Los Angeles County
Los Angeles, California 90057

Introduction

The purpose of this paper is not to argue whether or not demineralization of wastewater will some day be practiced, but to review some relevant research findings and cost estimates which may influence when and where such demineralization will occur. There are, of course, large quantities of wastewater in California with dissolved solids concentrations well above what might generally be conceded as acceptable limits. Such waters will probably not be reused unless they have been demineralized. However, there are many more situations in which wastewaters have marginal dissolved solids concentrations of slightly above or below 1,000 mg/l. In discussing waters of this quality with water supply agencies there appears to be considerable interest in reducing the dissolved solids to more desirable levels. This situation, which can be described as removing the increment of dissolved solids added to the initial water supply by one-time use in the community, seems to be the most probable application for demineralization in the near future, and most of the research work and cost estimates contained in this paper have been developed on that assumption.

Presented at the Workshop on Wastewater and Reuse sponsored by the University of California, Berkeley, held at South Lake Tahoe, Calif., June 25-26, 1970.

The Sanitation Districts have been involved in pilot scale evaluations of demineralization processes since 1965. The work has been conducted jointly with the Federal Water Quality Administration at the Sanitation Districts' Pomona Water Renovation Plant. A portion of the 10 mgd activated sludge plant effluent is diverted to the Research and Development Facility shown in Figure 1.* In addition to the various pilot plants, the research facility includes office and laboratory space for five engineers and seven chemists working on various studies. At the heart of this research facility is a 300,000 gallon per day, four-stage, downflow, packed-bed, granular activated carbon adsorption column. For the past five years this system has received typical secondary effluent from the Pomona plant and reliably produced a high quality carbon filtered water which has also served as the feed water for the demineralization processes. Average quality data for the activated sludge plant effluent and carbon adsorption product water are shown in Table 1.

Since all of the demineralization processes investigated at Pomona involve membranes or resins which are subject to fouling in the presence of high concentrations of organic matter, activated carbon adsorption is at present considered an integral part of these demineralization schemes. The cost of activated carbon treatment is estimated at 9¢ per thousand gallons for a 10 mgd facility, which must be added to the demineralization cost to arrive at the total cost estimates

*Omitted from this Seminar paper.

for each of the three processes presented at the end of this paper. The three processes -- reverse osmosis, electrodialysis, and ion exchange -- are presented in decreasing order of total estimated cost based on the studies to date.

Reverse Osmosis

Although the reverse osmosis process is the newest technological development of the three processes studied at Pomona, it was the first to be piloted there. There are many ways in which the principals of reverse osmosis can be converted to hardware and there are at least five or six companies with different versions of the process. Three of these have been piloted at Pomona but only two have been operated for sufficient length of time to permit a reasonable evaluation of their performance and economic potential. In the foreground of Figure 2* there is a 12,000 gallon per day unit manufactured by the Gulf General Atomic Corporation, and on the left edge of the slab a 3,600 gallon per day unit manufactured by Universal Water Corporation.

The Gulf General Atomic (GGA) unit features a spiral wound membrane packed into high pressure, 4" diameter pipes arranged in series. The configuration of the membrane module is illustrated in Figure 3. Each module is comprised of a wound up sandwich comprised of two layers of membrane for the "bread" around a "filling" of membrane support material. A brine side spacer separates layers of membrane as the membrane is wound around a low pressure product water tube at the central core. The brine stream passes longitudinally down the membrane through

* Omitted from this Seminar paper.

the spacer under an initial pressure of 465 psi. Water molecules pass through the membrane and travel spirally through the backing material to the low pressure product water collection tube at the center. Each module is 3' long and contains 50 square feet of membrane surface area. The unit shown in Figure 3 has a total surface area of 1,350 square feet and a flux rate of slightly under 10 gallons per day per square foot.

The Universal Water Corporation (UWC) unit features a more typical reverse osmosis design. The basic component of this system is a half-inch diameter perforated plastic tube which serves as the support for the membrane which is contained within the tube. Eighteen of these tubes are connected in series and housed within a 4" diameter product water collection pipe. The feed water flows through each tube at an initial pressure of 600 psi. The 4" diameter modules are connected in a combination of series and parallel arrangements to achieve the desired flow rate past each membrane. This unit contains 225 square feet of membrane area and has a flux rate of approximately 15 gallons per day per square foot, which is substantially greater than that of the GGA unit, even when accounting for the difference in operating pressure.

Although the present state of development of the reverse osmosis process results in the highest cost estimate of the three processes considered at Pomona, it is also the process with the

greatest potential for technological improvement. Significant improvements in either flux rate or membrane life could appreciably lower cost estimates. In order to provide reasonably reliable estimates for the present state of technology, the Sanitation Districts are operating both the GGA and UWC units on extended runs to determine the life of the membranes and the rate of change in flux rate with time. The performance of the two units is compared in Figure 4. At this time the GGA unit has been onstream about 9,000 hours, or approximately one year, compared to about 5,000 hours, or seven months, for the UWC unit. These runs are believed to be the longest continuous runs of a reverse osmosis system for the demineralization of reclaimed water. Even so, it will require another year of operation of the GGA unit to confirm the two-year life expectancy which has been assumed for purposes of the cost estimate on this process.

The difference in flux rate between the two units is substantial and apparently represents a difference in the state of technological development of membrane production by the two companies. The rate of decline in flux rate has been identical for each unit and has been empirically found to follow a straight line when plotted against time on a log-log scale, as shown in Figure 4. During the course of the life run the influent water has been automatically maintained at a pH of 5 with sulfuric acid. There is a daily flush with a mixture of air and tap water and a weekly flushing with an enzyme detergent. Whether

these routine operating procedures must be increased or can be reduced without affecting membrane life still needs to be determined.

Data on performance of the GGA unit is presented in Table 2. Average water recovery during this period was 75%, although it is anticipated that for a large installation a more advantageous flow pattern would permit recoveries of up to 85 percent. Because of the general interest in nutrient removal, Table 2 includes data on ammonia nitrogen, nitrate nitrogen and phosphate, as well as on TDS. It should be noted, however, that when these processes are used to remove the increment of TDS added during one municipal use of water, nutrient levels will not be reduced sufficiently to inhibit eutrophication problems. This process has maintained over 90% TDS reduction for an extended period and has the advantage of similarly high removal of COD materials. It is even more selective for the removal of phosphate but has only achieved about a 50% removal of nitrate nitrogen.

Electrodialysis

The 15,000 gallon per day electrodialysis unit constructed by Ionics, Inc. is shown in Figure 5* A rectifier in the gray chamber in the foreground provides a 70 volt DC output across the membrane stack, just visible on the left. The membrane stack is 18" x 20" x 12" deep and contains 65 anion and cation permeable membrane pairs and their respective spacers.

* Omitted from this Seminar paper.

The flow diagram is illustrated in Figure 6. A flow of approximately 12 gallons per minute in the feed stream is split with 10 or 11 gallons per minute passing through the dilute stream and becoming product water as the cations and anions migrate into the concentrate stream. The remaining one to two gallons per minute of the feed stream is mixed with 10 to 11 gallons per minute of brine recirculation flow to provide equal flow rates and consequently equal pressures on both sides of the membranes. A pH of 3.5 in the concentrate stream is maintained automatically by the injection of sulfuric acid.

The percent conductivity removal during the first three months of operation of this unit is presented in Figure 7. Decreases in percent conductivity removal are closely related to decreases in amperage across the unit. The unit is designed to operate at 4 amps and generally achieves a dissolved solids reduction of about 34 percent. On the average, conductivity removal has been relatively stable. Most of the fluctuations noted on the graph are related to the changing character of the TDS of the water rather than to changes in the membrane itself. Routine operating procedures to maintain high performance include flushing with a sequence of an enzyme detergent, tap water, and acid every second day with the voltage shut off. This has permitted at least four months of operation without requiring manual cleaning of the membranes. The relative cost of more frequent manual cleaning vs. routine flushing has not yet been determined. Performance data from a previous period of operation

is shown in Table 3. This process is more selective for nitrogen compounds and less selective for phosphate than for average TDS removal. This process has the disadvantage of being relatively ineffective on soluble organic material, as represented by the 15% removal of COD.

Ion Exchange

The ion exchange pilot plant shown in Figure 8* was originally supplied by Infilco Corporation with its own pretreatment unit. However, the pretreatment unit was unreliable and the cost of pretreatment was greater than the savings in ion exchange costs, which resulted in abandonment of the pretreatment portion and operation of the ion exchangers with carbon filtered water. A simplified flow diagram is presented in Figure 9 for this strong acid-weak base ion exchange process. The unit operates at 2.5 gallons per minute for a 16-hour period each day and produces about 2,400 gallons of product water per day. The primary cation exchanger contains 8 cubic feet of resin and the secondary cation exchanger 3.2 feet of resin, with each of the anion exchangers containing 1.6 cubic feet of resin. Once a day the last three exchangers are shut off and the primary cation exchanger is backwashed with feed water. Once a week each of the other exchangers is shut off in sequence and back-washed with its normal feed water.

A simplified diagram of the regeneration process is presented in Figure 10. Sulfuric acid is used to regenerate the cation

* Omitted from this condensed paper.

resins and ammonium hydroxide is used to regenerate the anion resins. The regenerates are discharged to waste and the resins are given a product water rinse.

Comparison of the feed and product water TDS for 80 days of operation is presented in Figure 11. The influent TDS averaged slightly over 600 mg/l, and the product water TDS averaged about 70 mg/l. Ion exchange systems are generally used in industry to produce extremely low concentrations of dissolved solids. For demineralization of reclaimed water, however, substantially higher TDS values are not only permissible but desired to reduce costs. By reducing the degree of regeneration of the resins and operating beyond the normal breakthrough point for industrial water quality, it has been possible to reduce the cost of regenerating the resins while producing a very acceptable demineralized product water.

By manipulating the end points, as previously described, the ion exchange system has accomplished an average TDS reduction of 87% at an average water recovery of 87%, as shown in Table 4. COD removal was much better than expected in this process and phosphate removal is outstanding, although this ion exchange system is less selective for ammonia nitrogen.

Cost Comparison

Tables 5, 6 and 7 present comparative cost estimates for the reverse osmosis, electrodialysis and ion exchange processes

based on producing 17 million gallons per day of product water. The reverse osmosis cost data in Table 6 still involves many assumptions. Firm fixed price cost has been provided by the Gulf General Atomic Corporation and is believed to be conservative, but until plants of this type are constructed on a large scale, reliable production costs will not be available. The cost of chemicals is comparatively high in this estimate and is largely represented by the cost of pH adjustment and enzyme detergent flushing. There may be some possibility of reducing this cost in the future. The item for "maintenance materials" is figured as 2% of the depreciable portion of the capital investment on all estimates and is a relatively minor portion of the total cost. Power is naturally a substantial figure for a reverse osmosis system and cannot be reduced. The largest item is membrane replacement which is one of the reasons membrane life and flux rate are the points where major emphasis on improving technology must be placed. The effect of changing either of these variables is illustrated in Figure 12. A two-year life and 10 gallons per day per square foot flux rate has been used in the previous cost estimate. Accomplishment of a three-year membrane life and an increase of flux rate to between 15 and 20 gallons per day per square foot would make a 1/3 reduction in the total cost of this process.

At first glance the estimate for electrodialysis in Table 6 looks relatively low. However, this cost, as opposed to that of reverse osmosis and ion exchange, does not have the potential

resins and ammonium hydroxide is used to regenerate the anion resins. The regenerates are discharged to waste and the resins are given a product water rinse.

Comparison of the feed and product water TDS for 80 days of operation is presented in Figure 11. The influent TDS averaged slightly over 600 mg/l, and the product water TDS averaged about 70 mg/l. Ion exchange systems are generally used in industry to produce extremely low concentrations of dissolved solids. For demineralization of reclaimed water, however, substantially higher TDS values are not only permissible but desired to reduce costs. By reducing the degree of regeneration of the resins and operating beyond the normal breakthrough point for industrial water quality, it has been possible to reduce the cost of regenerating the resins while producing a very acceptable demineralized product water.

By manipulating the end points, as previously described, the ion exchange system has accomplished an average TDS reduction of 87% at an average water recovery of 87%, as shown in Table 4. COD removal was much better than expected in this process and phosphate removal is outstanding, although this ion exchange system is less selective for ammonia nitrogen.

Cost Comparison

Tables 5, 6 and 7 present comparative cost estimates for the reverse osmosis, electrodialysis and ion exchange processes

based on product of 10 million gallons per day of product water. The reverse osmosis cost [illegible] still involves many assumptions. Part [illegible] cost has been provided by the Gulf General Atomic Corporation and is believed to be conservative, but until plants of this type are constructed on a large scale, reliable production costs will not be available. The cost of chemicals is surprisingly high in this estimate and is largely represented by the cost of pH adjustment and enzyme detergent flushing. There may be some possibility of reducing this cost in the future. The item for "maintenance materials" is figured as 1% of the depreciable portion of the capital investment on all estimates and is a relatively minor portion of the total cost. Power is naturally a substantial figure for a reverse osmosis system and cannot be reduced. The largest item is membrane replacement which is one of the reasons membrane life and flux rate are the points where major emphasis on improving technology must be placed. The effect of changing either of these variables is illustrated in Figure 12. A two-year life and 10 gallons per day per square foot flux rate has been used in the previous cost estimate. Accomplishment of a three-year membrane life and an increase of flux rate to between 15 and 20 gallons per day per square foot would make a 1/3 reduction in the total cost of this process.

At first glance the estimate for electrodialysis in Table 6 looks relatively low. However, this cost, as opposed to that of reverse osmosis and ion exchange, does not have the potential

resins and ammonium hydroxide is used to regenerate the anion resins. The regenerates are discharged to waste and the resins are given a product water rinse.

Comparison of the feed and product water TDS for 80 days of operation is presented in Figure 11. The influent TDS averaged slightly over 600 mg/l, and the product water TDS averaged about 70 mg/l. Ion exchange systems are generally used in industry to produce extremely low concentrations of dissolved solids. For demineralization of reclaimed water, however, substantially higher TDS values are not only permissible but desired to reduce costs. By reducing the degree of regeneration of the resins and operating beyond the normal breakthrough point for industrial water quality, it has been possible to reduce the cost of regenerating the resins while producing a very acceptable demineralized product water.

By manipulating the end points, as previously described, the ion exchange system has accomplished an average TDS reduction of 87% at an average water recovery of 87%, as shown in Table 4. COD removal was much better than expected in this process and phosphate removal is outstanding, although this ion exchange system is less selective for ammonia nitrogen.

Cost Comparison

Tables 5, 6 and 7 present comparative cost estimates for the reverse osmosis, electrodialysis and ion exchange processes

of being reduced by blending since the process only achieves about a 1/3 reduction in TDS. Furthermore, it is felt that the capital cost of the process as well as the operating cost are relatively well known, with only minor possibilities of making slight improvements in either chemical usage or labor costs in the future.

The cost of ion exchange, as presented in Table 7, appears to be the most reliable estimate of the three processes, although the cost of regenerants is related to the particular combination of ions included in the overall dissolved solids of the feed water. Furthermore, this process is the most sensitive to waters of variable or high dissolved solids content because the cost of regenerants is directly related to the number of ions removed. With regeneration costs amounting to 16.2¢ per thousand gallons out of the total estimated cost of 22¢ per thousand gallons, it is apparent why most of the effort has centered on improving the efficiency of utilization of regenerants Resin replacement is not a particularly large cost item, although the 20% per year figure is considered somewhat conservative.

In Figure 13 the total cost of the three processes are compared under equivalent conditions for an influent TDS range from 500 to 1,500 mg/l. In each case it has been assumed the influent TDS is reduced by 33% through demineralization. For the electrodialysis unit 33% is approximately the reduction obtained in a single pass. However, both the reverse osmosis and ion

exchange units provide 90% TDS removal, hence the cost of these processes have been adjusted by assuming a mixture of demineralized and non-demineralized carbon effluent to produce 20 mgd of the desired TDS product concentration. Though these estimates include the 9¢ per thousand gallons cost of carbon filtration, they do not include the cost of brine disposal. On the coastal plain of Los Angeles County, brine disposal is not a problem since brines can be returned to the sewerage system and ultimately reach the Pacific Ocean. For inland plants this problem would be of such major proportions as to essentially eliminate consideration of demineralization of reclaimed wastewaters in such locations. Throughout the range of TDS considered, the ion exchange process remains the most economical and amounts to about 20¢ per thousand gallons for a 33% TDS reduction for an influent TDS of 1,000 mg/l. The ion exchange cost curves have therefore been the basis for the generalized cost curves presented in Figures 14 and 15.

Figure 14 shows the estimated cost to obtain a particular product water TDS for a range of influent TDS values up to 1,500 mg/l. Figure 15 shows the estimated cost of achieving a selected amount of TDS removal for the same influent TDS range. These costs include pretreatment by carbon adsorption and exclude the cost of brine disposal. The horizontal plots in Figure 15 illustrate the nature of the ion exchange process since the cost of the process is essentially proportional to the ions removed or exchanged, regardless of the influent TDS.

All of the previous estimates have been based on the cost of a 10 mgd demineralization facility. An estimate for a 10 mgd facility is obviously quite an extrapolation from the 2,000 to 15,000 gallons per day pilot plant units which have actually been tested. In Figure 16 the costs of achieving several different product water TDS concentrations with an influent TDS of 1,000 mg/l have been estimated for plant sizes from 0.5 to 10 mgd based on the ion exchange process. It appears that extremely small plant sizes will remain uneconomical but that there may be possibilities for minor improvements in cost for plants larger than 10 mgd. A recent article in the Journal of the American Water Works Association by William R. Gianelli (April, 1970) includes an estimate of 37¢ per thousand gallons for desalting sea water in a 50 mgd plant combined with a large nuclear power generating facility. With no knowledge of what assumptions were actually used in the estimate, it is of interest to note that it is quite comparable to the cost of reducing the TDS in 10 mgd of secondary effluent from 1,000 mg/l to about 500 mg/l obtained from Figure 16.

Summary

The strong acid-weak base ion exchange process piloted at Pomona has proven to be more economical and reliable than the reverse osmosis or electrodialysis systems with which it has been compared. Practical reductions in dissolved solids can be obtained for costs of 20 to 30 cents per thousand gallons where brines can be discharged to the ocean. These costs

should be confirmed by construction of a larger pilot facility but indicate that reclamation, demineralization and reuse of wastewater should be considered prior to demineralization of seawater and carefully compared with the estimated costs of future imported supplies.

Credits

The research work reported upon in this paper was jointly financed by the Federal Water Quality Administration and the Sanitation Districts of Los Angeles County. The following project engineers from both agencies have operated the pilot units, collected and analyzed data, and overcame numerous problems in developing the information upon which this paper is based:

Project Engineers	Process
Max Augustus	Reverse Osmosis
Robert Miele	Reverse Osmosis
James Gratteau	Reverse Osmosis
John English	Electrodialysis
John Burgeson	Electrodialysis
Harold Takenaka	Electrodialysis
Hans Krock	Ion Exchange
Ching-lin Chen	Ion Exchange

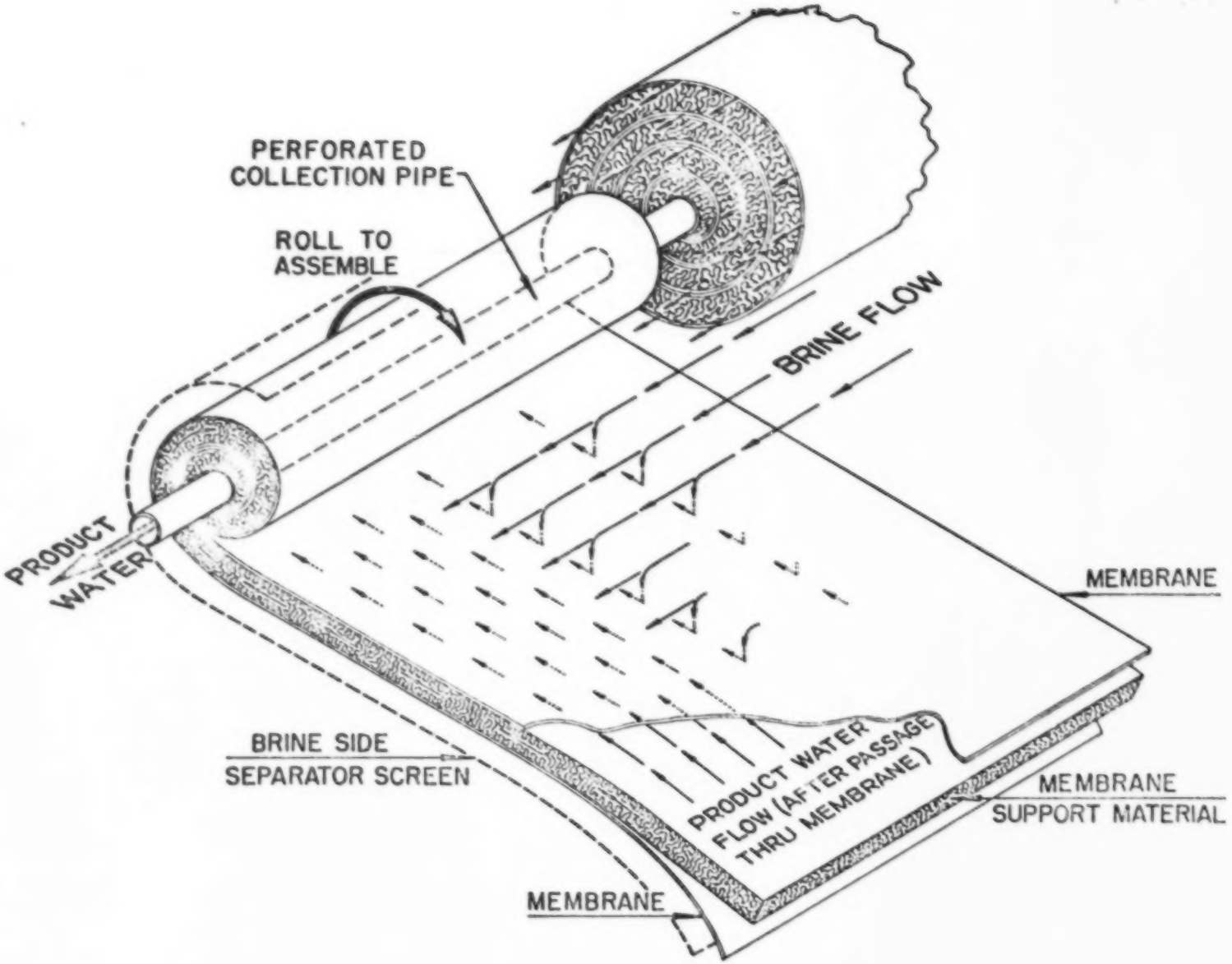


FIGURE 3: SCHEMATIC OF SPIRAL WOUND REVERSE OSMOSIS MODULE

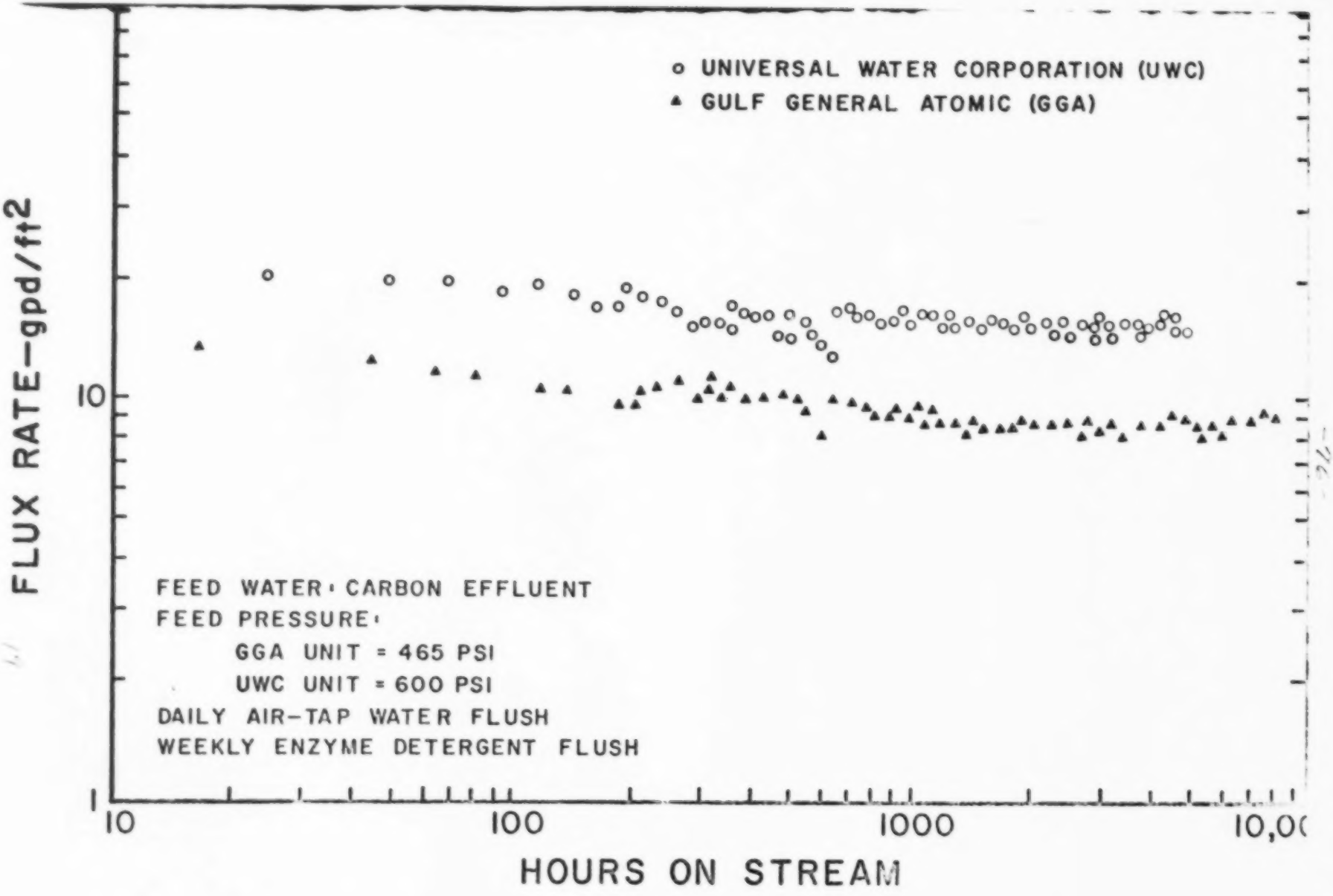


FIGURE 4: PERFORMANCE OF REVERSE OSMOSIS UNITS

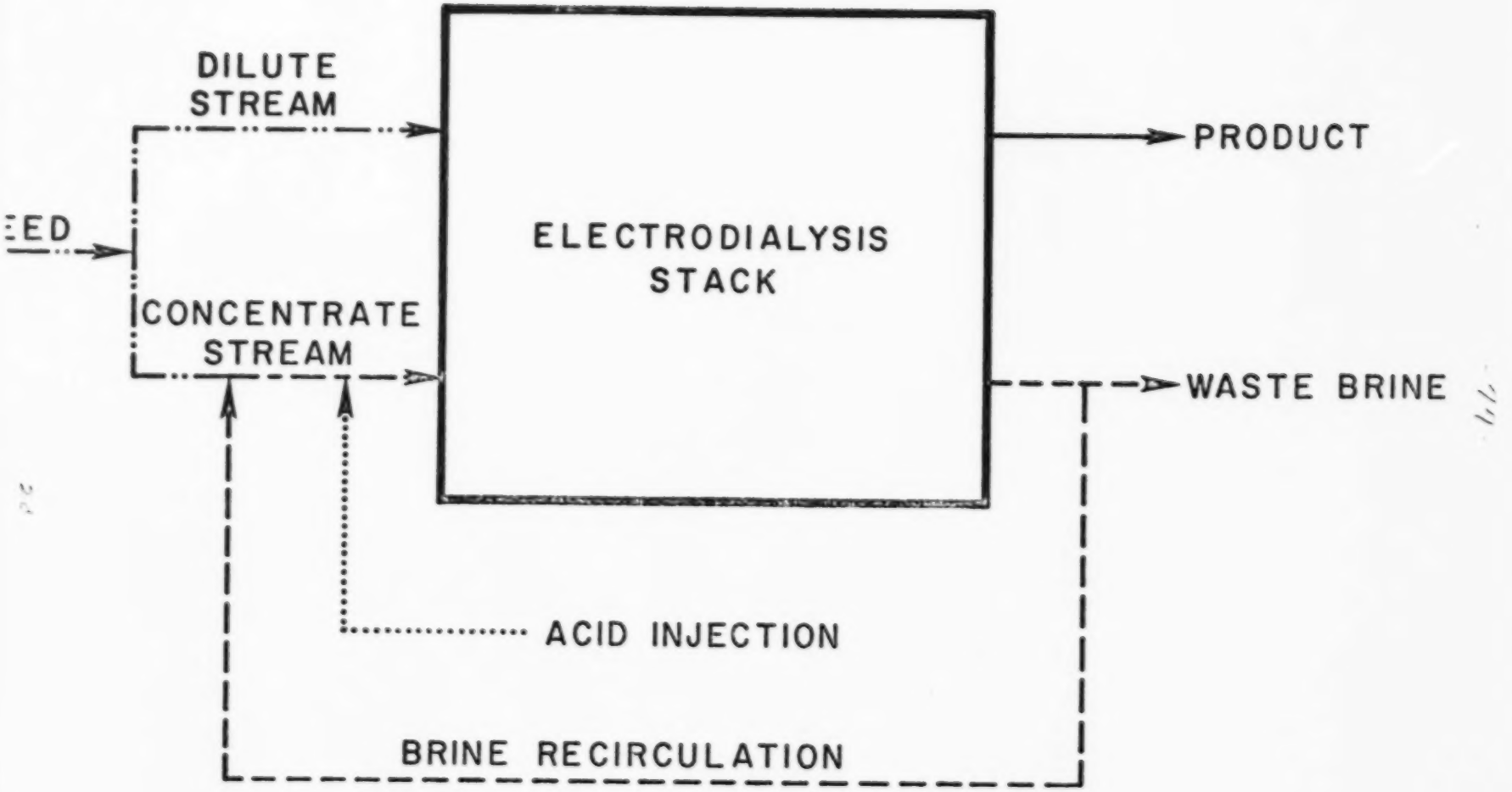


FIGURE 6: SCHEMATIC FLOW DIAGRAM OF ELECTRODIALYSIS PILOT PLANT

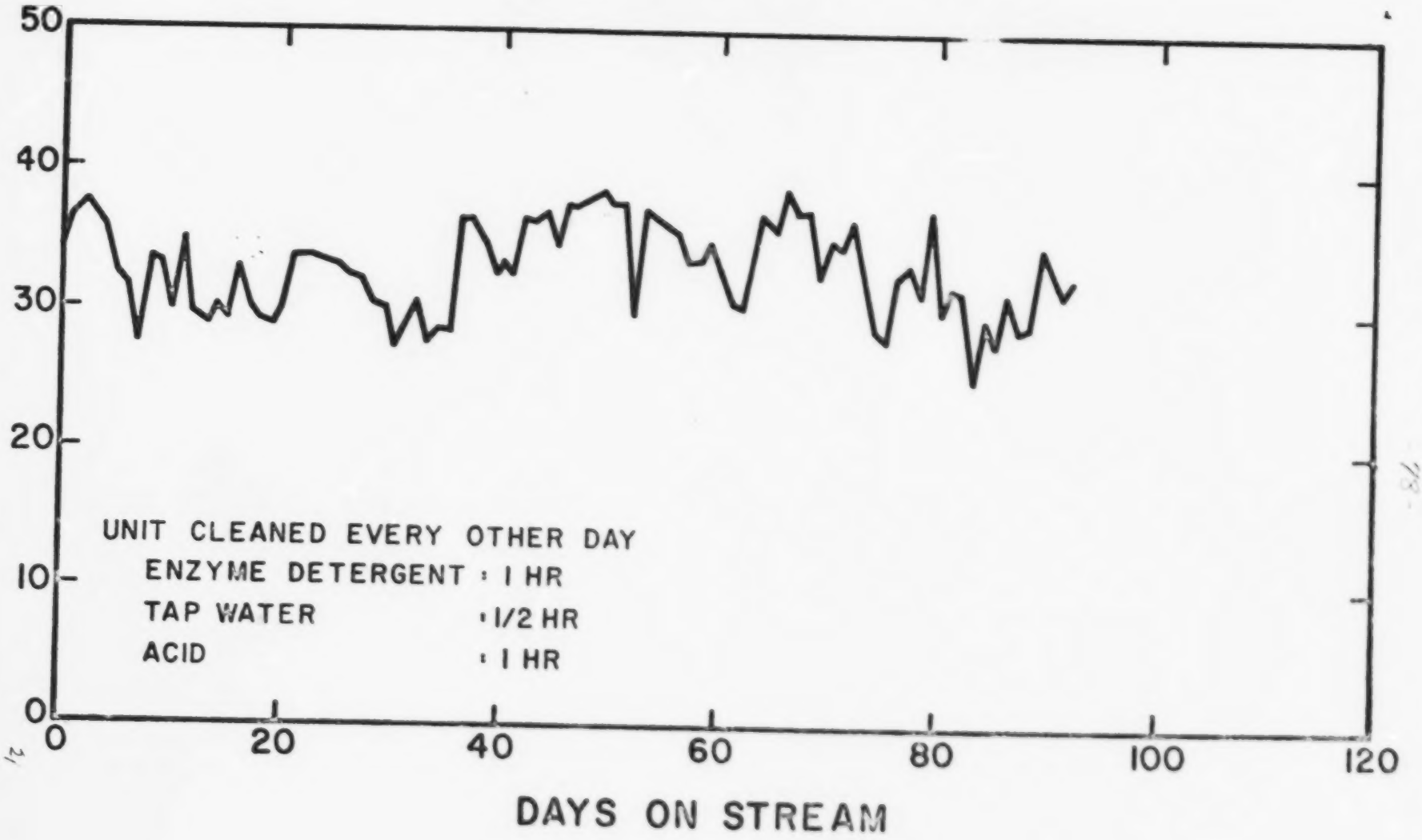


FIGURE 7: PERFORMANCE OF ELECTRODIALYSIS UNIT

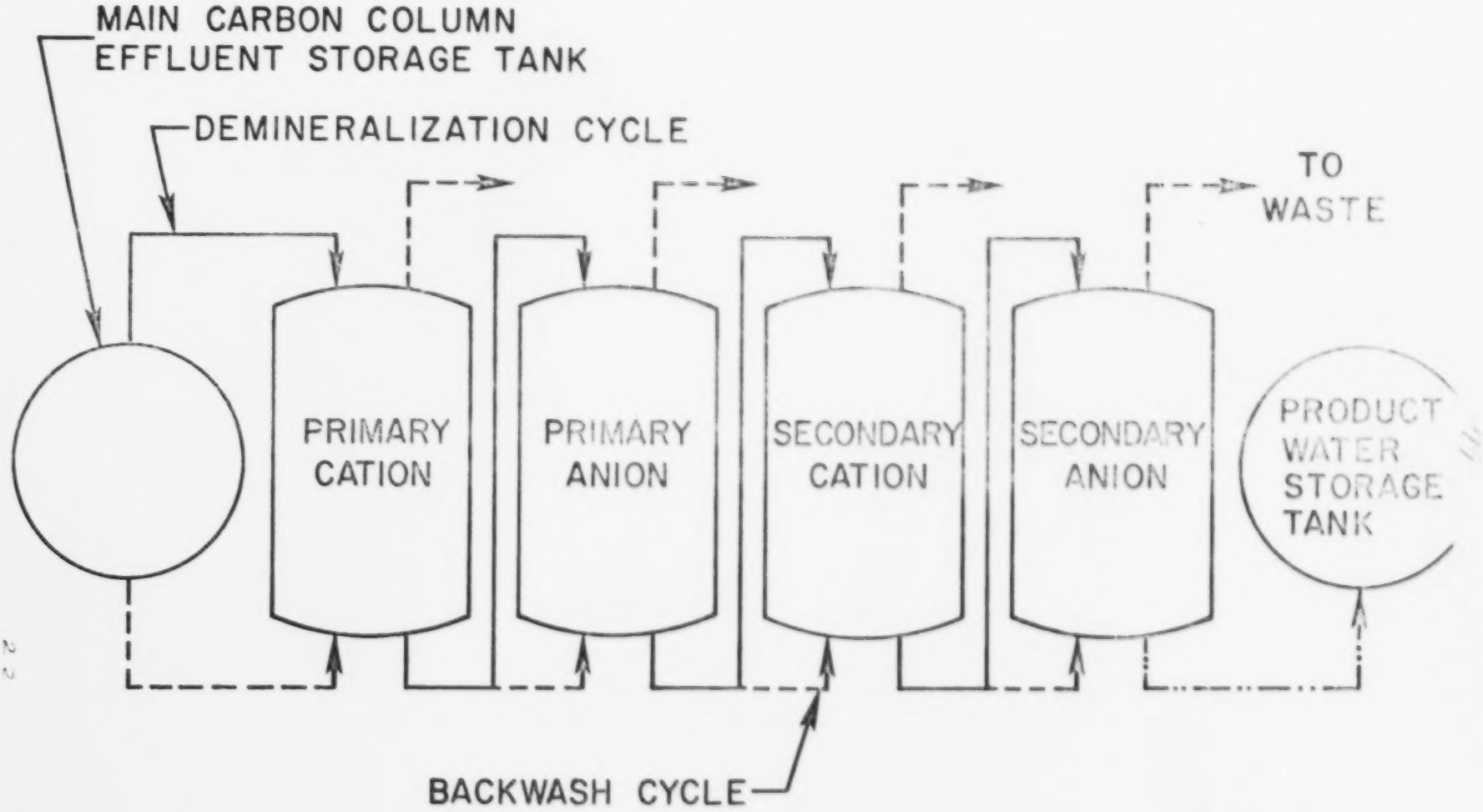


FIGURE 9: SCHEMATIC FLOW DIAGRAM OF ION EXCHANGE UNIT

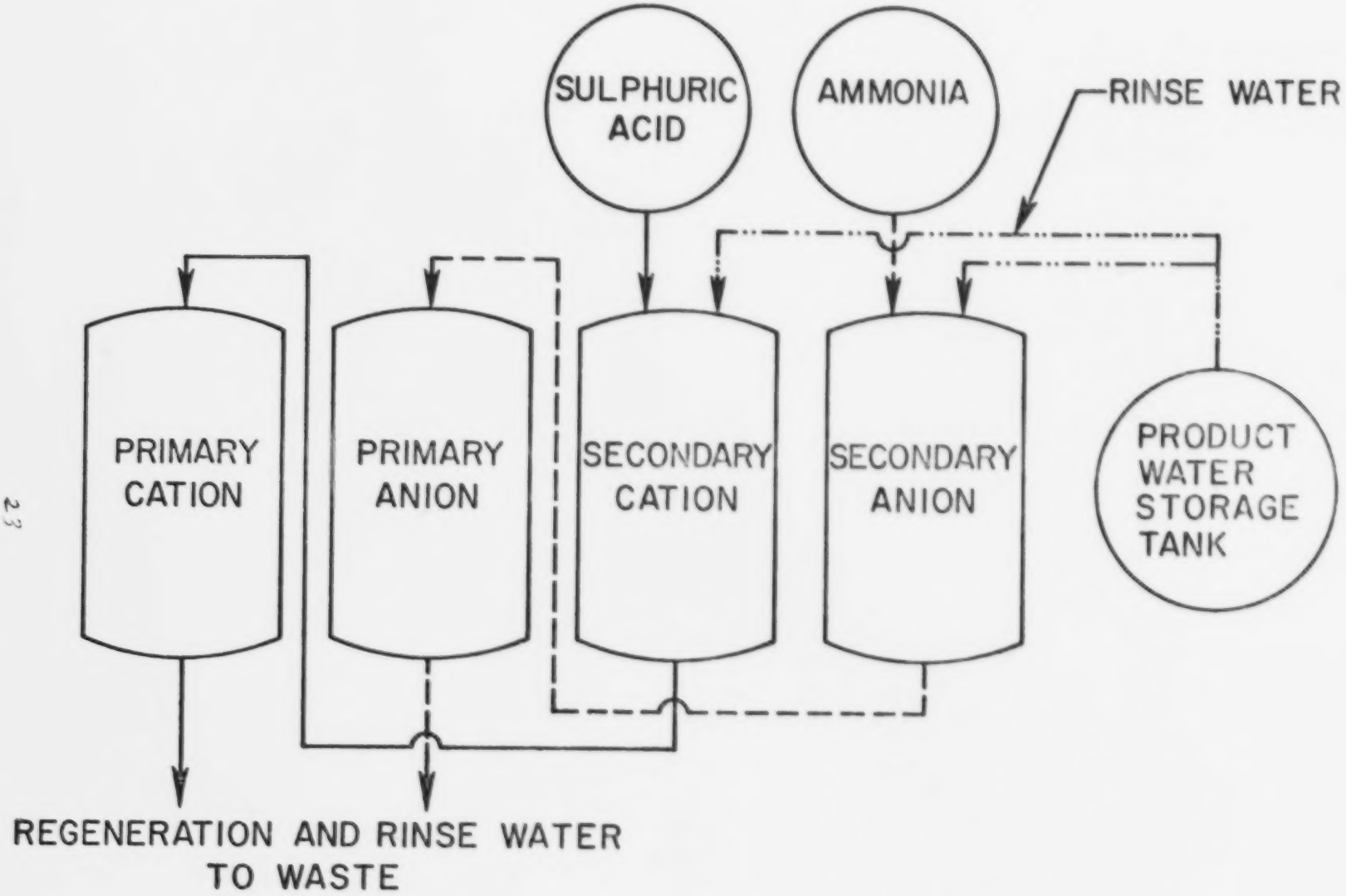


FIGURE 10: SCHEMATIC OF ION EXCHANGE REGENERATION PROCESS

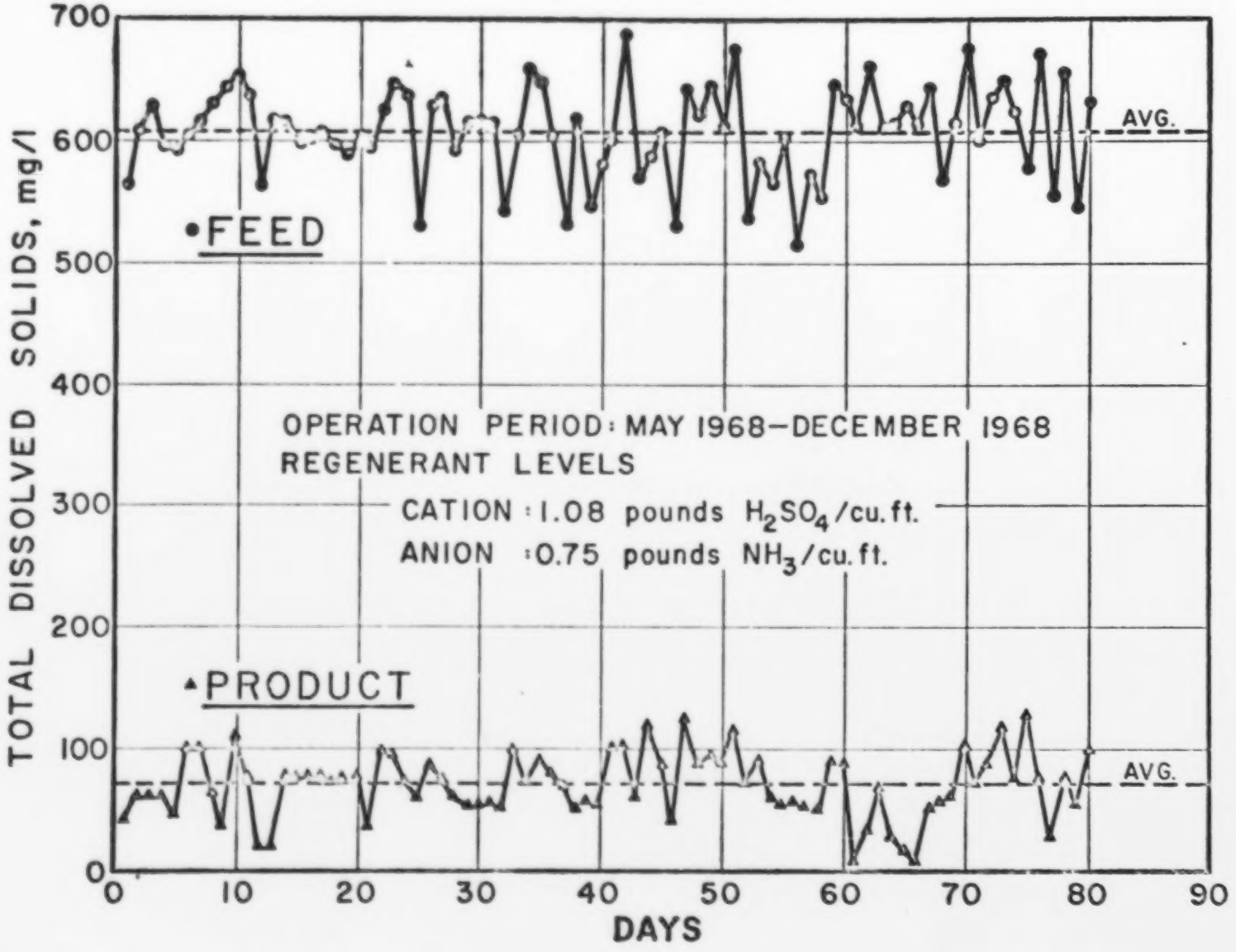
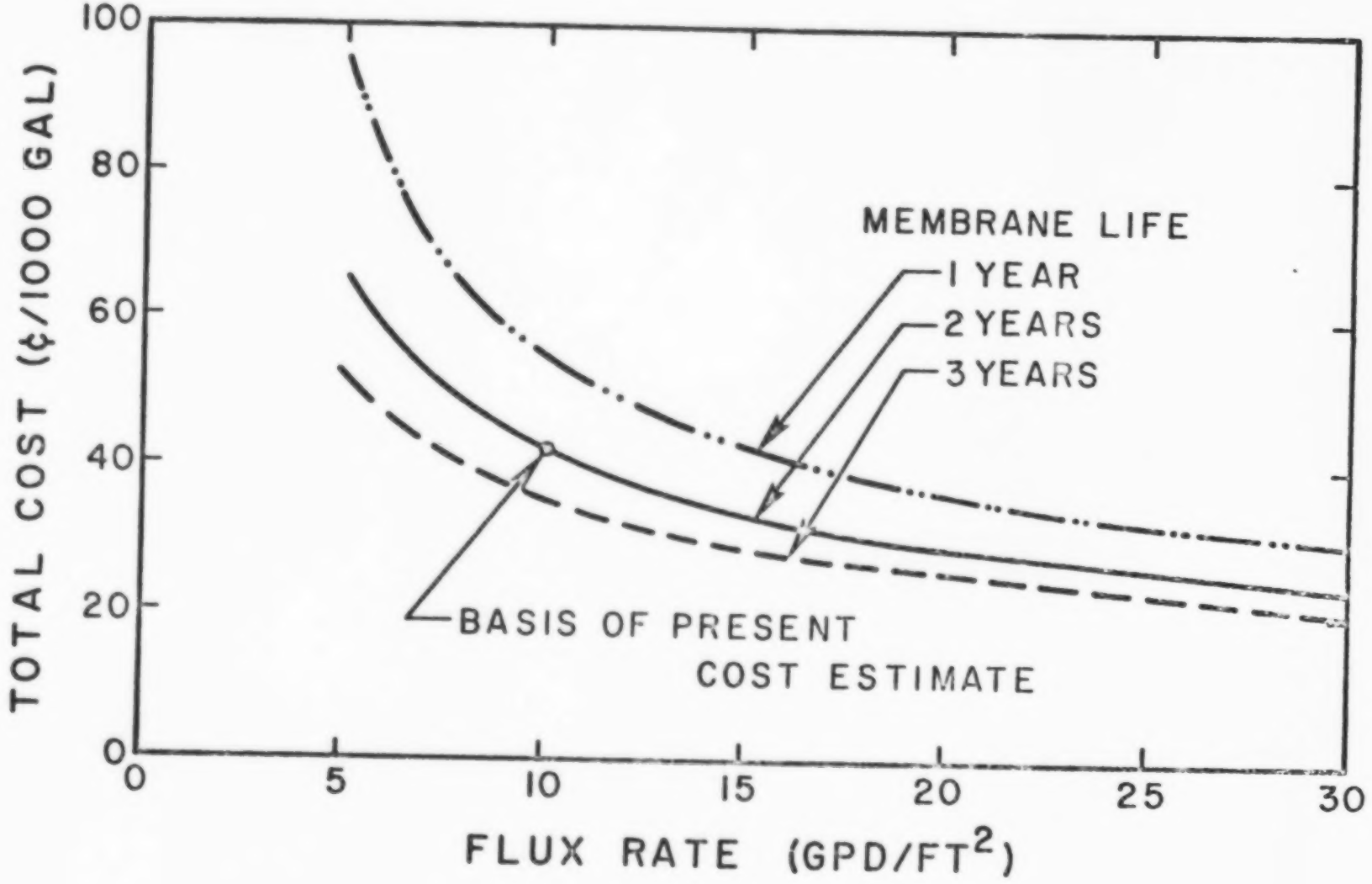


FIGURE 11: PERFORMANCE OF ION EXCHANGE UNIT


TOTAL COST (¢/1000 GAL)
100
80
60
40
20
0
0
5
10
15
20
25
30
FLUX RATE (GPD/FT²)
MEMBRANE LIFE
1 YEAR
2 YEARS
3 YEARS
BASIS OF PRESENT
COST ESTIMATE

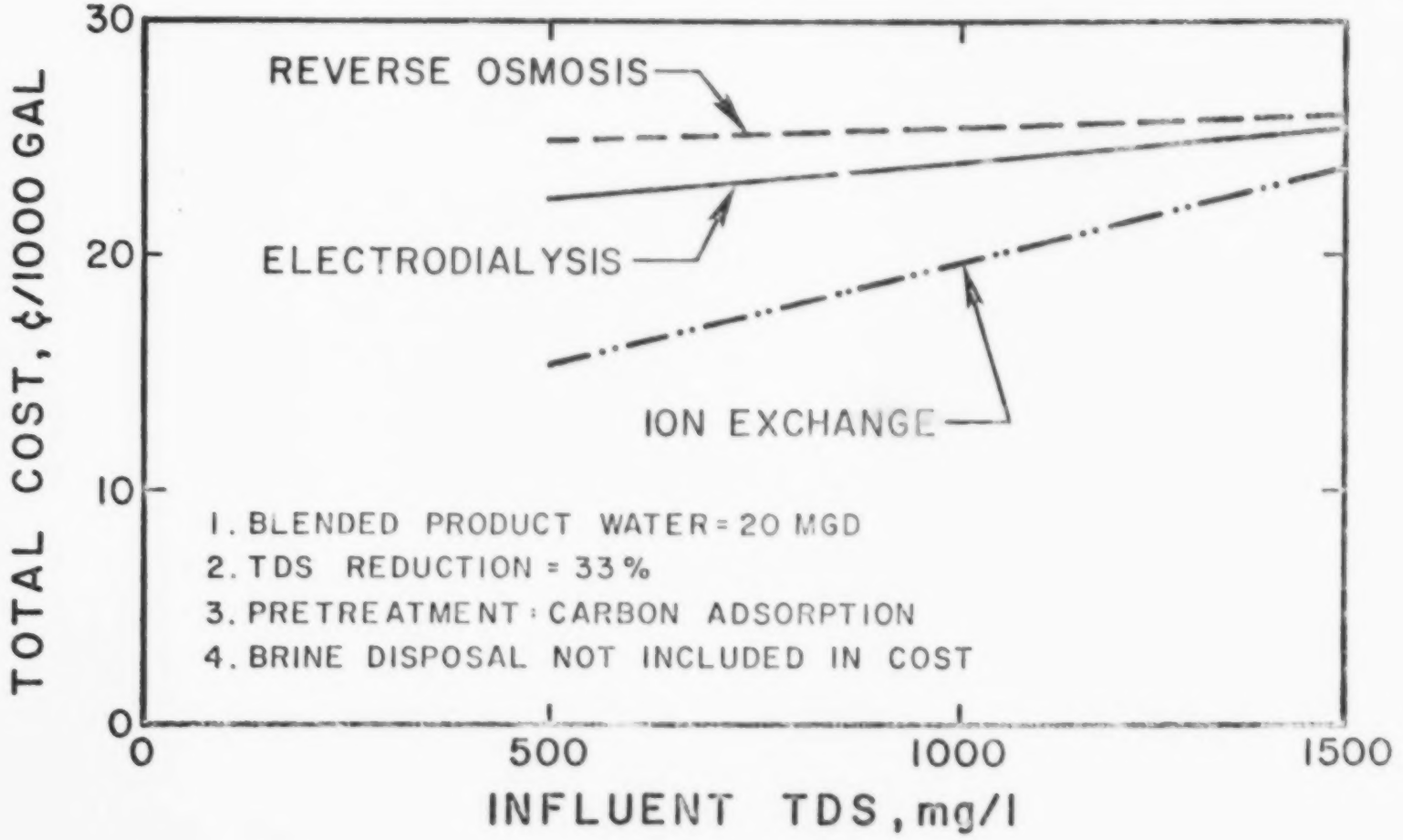


FIGURE 13: COMPARISON OF DEMINERALIZATION COSTS

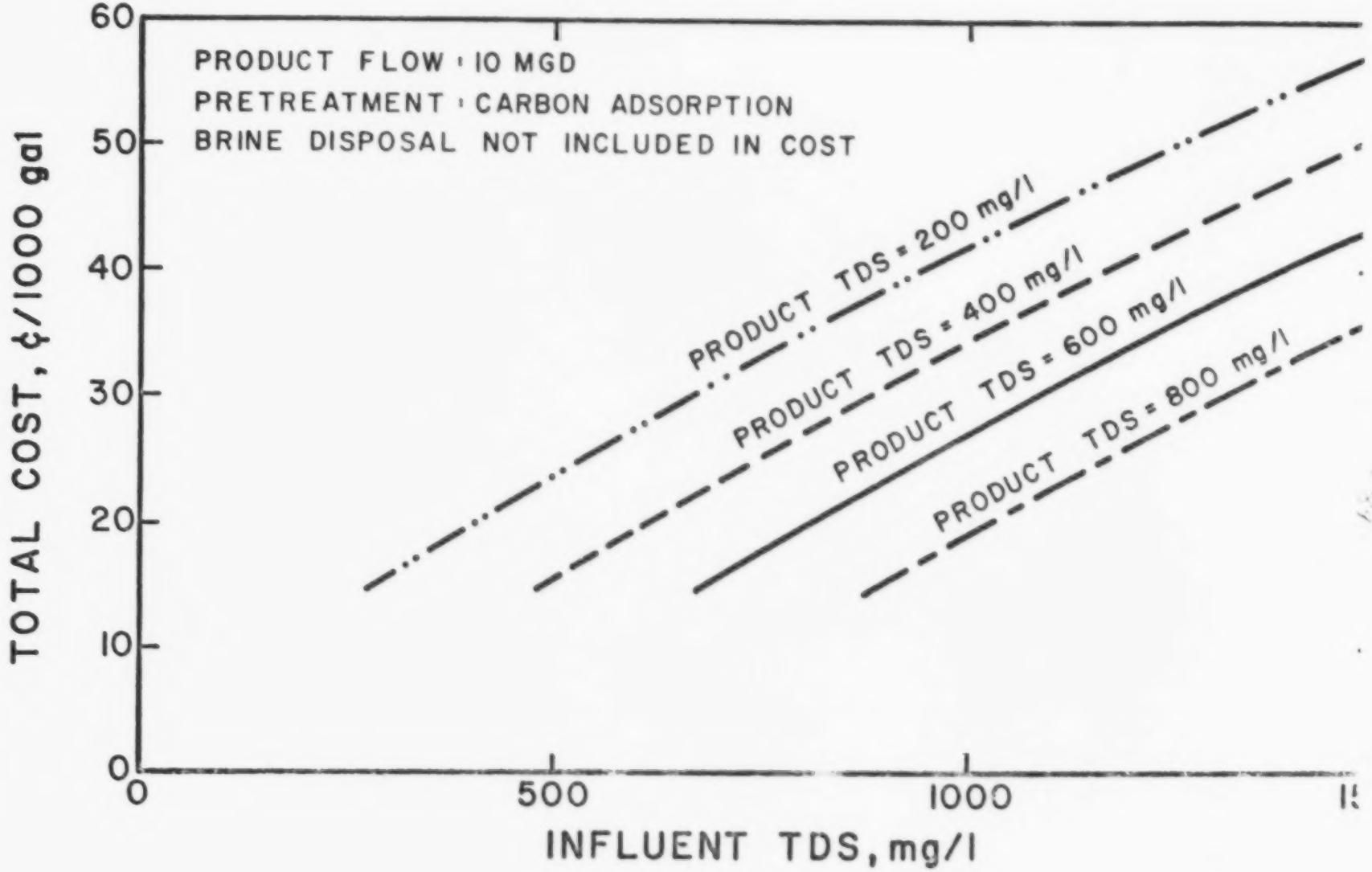


FIGURE 14: COST TO OBTAIN DESIRED PRODUCT TDS

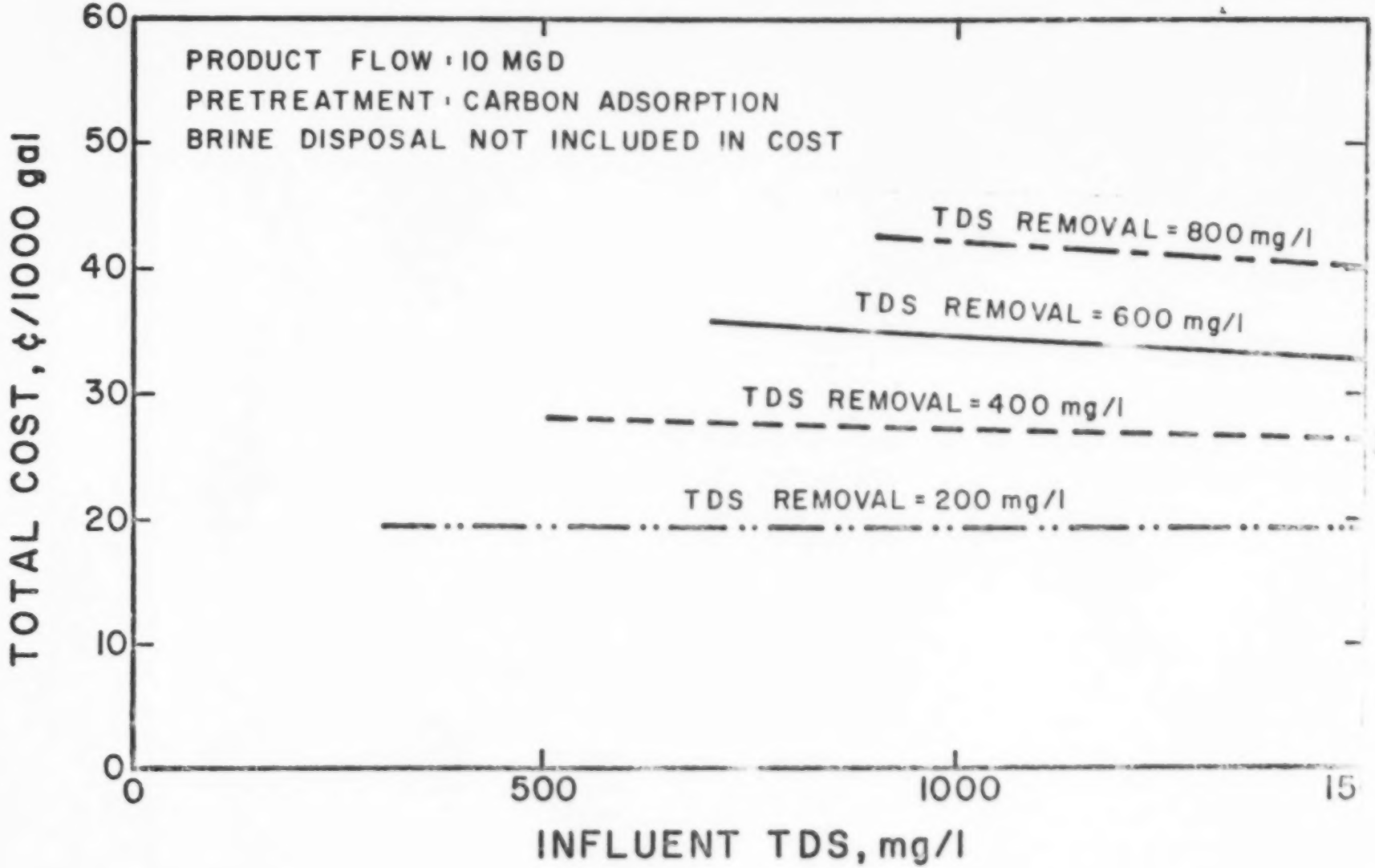


FIGURE 15: COST TO ACHIEVE DESIRED TDS REMOVAL

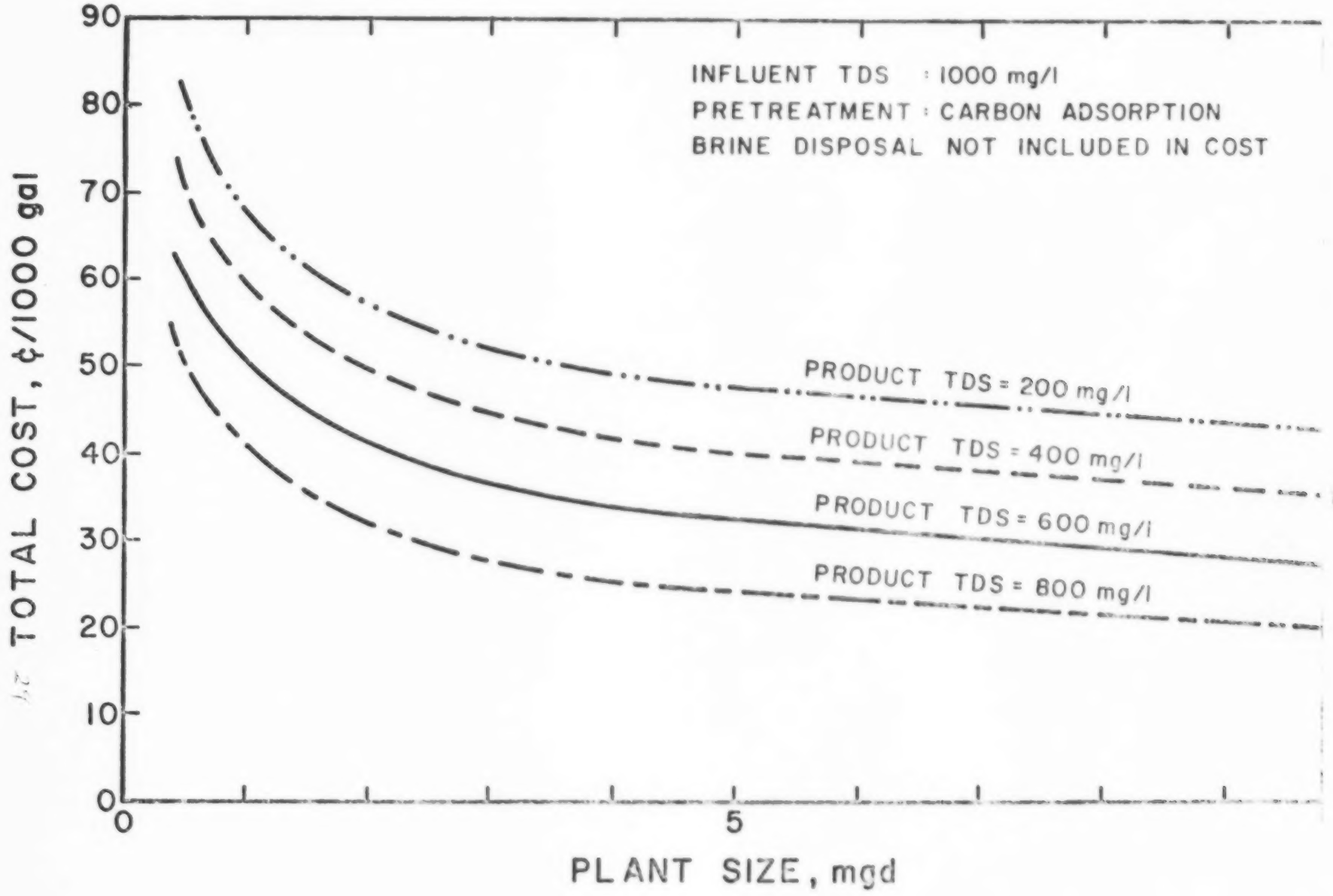


FIGURE 16: EFFECT OF PLANT SIZE UPON COST

PARAMETER		INFLUENT	EFFLUENT
SUSPENDED SOLIDS	mg/l	9	0.6
COD	mg/l	43	10
DISSOLVED COD	mg/l	30	8
TOC	mg/l	12	3
NITRATE, as N	mg/l	8.1	6.6
TURBIDITY	JTU	8.2	1.2
COLOR		28	3
ODOR		12	1
CCE	mg/l	—	0.026

TABLE 1: AVERAGE QUALITY DATA FOR CARBON ADSORPTION PILOT PLANT

TEST (mg/l)	INFLUENT	EFFLUENT	% REMOVA
COD	8.7	1.0	88.5
NH_3-N	10.1	1.1	89.2
NO_3-N	4.9	2.4	51.0
PO_4-P	10.9	0.2	98.4
TDS	750	59	92.1

NOTES

1. 15,000 GPD PILOT PLANT
2. INFLUENT IS CARBON EFFLUENT
3. AVERAGES BASED ON 40 SAMPLES (JUNE-SEPT. 1969)
4. ANALYSES RUN ON GRAB SAMPLES
5. AVERAGE WATER RECOVERY = 75%

TABLE 2: WATER QUALITY CHARACTERISTICS OF REVERSE OSMOSIS PROCESS

TEST (mg/l)	INFLUENT	EFFLUENT	% REMOVAL
COD	9.4	8.0	14.9
NH_3-N	9.0	5.1	43.4
NO_3-N	6.2	3.1	50.0
PO_4-P	10.1	7.8	22.8
TDS	705	465	34.0

NOTES

1. 15,000 GPD PILOT PLANT
2. INFLUENT IS CARBON EFFLUENT
3. AVERAGES BASED ON 10 SAMPLES (JUNE-AUG 1969)
4. ANALYSES RUN ON 24-HOUR COMPOSITE SAMPLES
5. AVERAGE WATER RECOVERY = 83%

TABLE 3: WATER QUALITY CHARACTERISTICS OF ELECTRODIALYSIS PROCESS

TEST (mg/l)	INFLUENT	EFFLUENT	% REMOVA
COD	10.0	3.7	63.0
NH_3-N	15.6	3.8	75.6
NO_3-N	2.9	0.4	86.2
PO_4-P	8.8	0.1	98.8
TDS	611	81	86.7

NOTES

1. 5000 GPD PILOT PLANT
2. INFLUENT IS CARBON EFFLUENT
3. AVERAGES BASED ON 20 SAMPLES (MAY-DEC 1968)
4. ANALYSES RUN ON 16-HOUR COMPOSITE SAMPLES
5. AVERAGE WATER RECOVERY = 87%

TABLE 4: WATER QUALITY CHARACTERISTICS OF ION EXCHANGE PROCESS

AMORTIZATION OF CAPITAL		¢/1000 GALLONS
$4,020,000; 20 YEARS @ 5%		8.8
OPERATION AND MAINTENANCE		
CHEMICALS	7.3	
MEMBRANE REPLACEMENT	14.4	
MAINTENANCE MATERIALS	2.6	
POWER	7.0	
LABOR	1.5	
		32.8
TOTAL		41.6

ASSUMPTIONS

1. INFLUENT TDS = 700 mg/l; EFFLUENT TDS = 100 mg/l
2. MEMBRANES: FLUX RATE = 10 GPD/FT^2; LIFE = 2 YEARS
3. WATER RECOVERY = 85%

TABLE 5: COST ESTIMATE FOR 10 MGD REVERSE OSMOSIS PLANT

AMORTIZATION OF CAPITAL	¢/1000 GALLONS	
$1,830,000 ; 20 YEARS @ 5%		4.0
OPERATION AND MAINTENANCE		
CHEMICALS	4.4	
MEMBRANE REPLACEMENT	3.3	
MAINTENANCE MATERIALS	0.9	
POWER	1.9	
LABOR	2.5	
		13.0
TOTAL		17.0

ASSUMPTIONS

1. INFLUENT TDS = 700 mg/l ; EFFLUENT TDS = 450 mg/l
2. MEMBRANE LIFE = 5 YEARS
3. WATER RECOVERY = 90 %

TABLE 6: COST ESTIMATE FOR 10 MGD ELECTRODIALYSIS PLANT

AMORTIZATION OF CAPITAL	¢/1000 GALLONS	
$1,370,000 ; 20 YEARS @ 5%		3.0
OPERATION AND MAINTENANCE		
REGENERATION		
CATION (SULFURIC ACID)	10.0	
ANION (AMMONIA)	4.2	
RESIN REPLACEMENT	2.0	
MAINTENANCE MATERIALS	0.6	
POWER	0.8	
LABOR	1.4	
		19.0
TOTAL		22.0

ASSUMPTIONS

1. INFLUENT TDS = 700 mg/l ; EFFLUENT TDS = 75 mg/l
2. REGENERATION EFFICIENCY : CATION = 85% ; ANION = 90%
3. RESIN REPLACEMENT : CATION = 10% AND ANION = 20% PER YEA
4. WATER RECOVERY = 87%